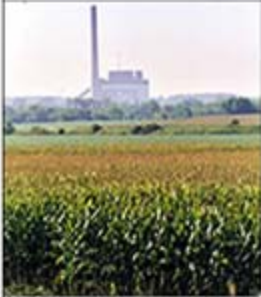




WESTERN GOVERNORS' ASSOCIATION



Strategic Assessment of Bioenergy Development in the West

Bioenergy Conversion Technology Characteristics

FINAL REPORT

Antares Group Inc.
September 1, 2008



Strategic Assessment of Bioenergy Development in the West

In February 2008, the Western Governors' Association adopted a policy reaffirming the governors' strong commitment to enhance and diversify the region's transportation fuels portfolio. The *Strategic Assessment of Bioenergy Development in the West* represents a major step in fulfilling that commitment and expands upon earlier work through WGA's Clean and Diversified Energy Initiative and the Transportation Fuels Initiative. The Bioenergy Assessment Team was formed to examine the potential for future development and to create a comprehensive framework to assess environmental, technical and socioeconomic impacts associated with national, state and regional bioenergy and biomass management policies. This assessment will assist the governors individually and collectively as they develop bioenergy policies. The extensive evaluations conducted by the Assessment Team are contained in the following areas:

- Biomass Resources in the Western States
- Biofuel Conversion Technologies
- Spatial Analysis and Supply Curve Development
- Analyses of Deployment Scenarios and Policy Interactions

While the assessment represents the consensus view of its authors, it is not adopted policy and does not represent the views of WGA or any individual Western Governor. Support for this work was provided by the U.S. Department of Agriculture.

Members of the Western Bioenergy Assessment Team

Western Governors' Association: Gayle Gordon

University of California, Davis: Nathan Parker, Peter Tittmann, Quinn Hart, Mui Lay, Joshua Cunningham and Bryan Jenkins

Kansas State University: Richard Nelson

U.S. Forest Service: Ken Skog, Marcia Patton Mallory, Robert Rummer and R. James Barbour

U.S. Department of the Interior: John Stewart

Antares Group, Inc.: Edward Gray, Anneliese Schmidt and Chris Lindsey

Future Resources Association: Gregg Morris

National Renewable Energy Laboratory: Anelia Milbrandt

Consultants: Jim Kerstetter and Ralph Overend

TABLE OF CONTENTS

| | | |
|----------|---|-----------|
| 1 | INTRODUCTION..... | 1 |
| 2 | CURRENT TECHNOLOGIES..... | 4 |
| 2.1 | GRAIN TO ETHANOL CONVERSION..... | 4 |
| 2.1.1 | <i>Technology Description.....</i> | 4 |
| 2.1.2 | <i>Performance and cost analysis.....</i> | 6 |
| 2.2 | FATTY ACID TO METHYL ESTERS: TRANSESTERIFICATION (BIODIESEL) | 10 |
| 2.2.1 | <i>Technology Description.....</i> | 10 |
| 2.2.2 | <i>Performance and cost analysis.....</i> | 12 |
| 2.3 | SUGAR TO ETHANOL FERMENTATION | 14 |
| 3 | TECHNOLOGIES READY FOR COMMERCIALIZATION IN THE MIDTERM (2010 – 2025)..... | 16 |
| 3.1 | LIGNOCELLULOSICS TO ETHANOL: HYDROLYSIS AND FERMENTATION | 16 |
| 3.1.1 | <i>Technology Description.....</i> | 17 |
| 3.1.2 | <i>Performance and cost analysis.....</i> | 20 |
| 3.2 | LIGNOCELLULOSICS TO ETHANOL: THERMOCHEMICAL CONVERSION AND FERMENTATION | 27 |
| 3.2.1 | <i>Technology Description.....</i> | 28 |
| 3.3 | LIGNOCELLULOSICS TO MIDDLE DISTILLATES: FISCHER TROPSCH | 30 |
| 3.3.1 | <i>Technology Description.....</i> | 30 |
| 3.3.2 | <i>Performance and cost analysis.....</i> | 32 |
| 3.4 | LIGNOCELLULOSICS TO MIXED ALCOHOLS: THERMOCHEMICAL CONVERSION..... | 36 |
| 3.4.1 | <i>Technology Description.....</i> | 36 |
| 3.4.2 | <i>Performance and cost analysis.....</i> | 38 |
| 3.5 | LIGNOCELLULOSICS TO GASOLINE: UPGRADING PYROLYSIS OIL | 40 |
| 3.5.1 | <i>Technology Description.....</i> | 40 |
| 3.5.2 | <i>Performance and cost analysis.....</i> | 44 |
| 3.6 | FATTY ACIDS TO HYDROCARBON (RENEWABLE DIESEL): HYDROTREATMENT | 47 |
| 3.6.1 | <i>Technology Description.....</i> | 48 |
| 3.6.2 | <i>Performance and cost analysis.....</i> | 50 |
| 4 | EMISSIONS DATA FOR BIOFUEL CONVERSION TECHNOLOGIES | 53 |
| 4.1 | YIELDS AND WATER USE..... | 53 |
| 4.2 | GREENHOUSE GAS AND CRITERIA POLLUTANT EMISSIONS..... | 54 |
| | REFERENCES..... | 58 |
| | APPENDIX A CONVERSION TECHNOLOGY STATUS MATRIX | 62 |
| | APPENDIX B CONVERSION TECHNOLOGY PRETREATMENT MATRIX..... | 64 |
| | APPENDIX C ECONOMIC ANALYSIS FACTORS | 65 |
| | APPENDIX D ADDITIONAL INFORMATION - GRAIN TO ETHANOL PROCESSES | 67 |
| | APPENDIX E OILSEED CROP FEEDSTOCK CONVERSION TO OIL | 70 |
| | APPENDIX F BIOMASS RECOVERED FROM MSW..... | 72 |
| | APPENDIX G CONVERSION TECHNOLOGY MODEL CALCULATORS..... | 73 |

List of Figures

| | |
|---|----|
| FIGURE 1 – SCHEMATIC DIAGRAM OF DRY MILLING ETHANOL PRODUCTION PROCESS | 4 |
| FIGURE 2 – SCHEMATIC DIAGRAM OF WET-MILLING PROCESS FOR CONVERSION OF GRAIN TO STARCH | 6 |
| FIGURE 3 – PROJECTED CORN PRICES..... | 10 |
| FIGURE 4 – SCHEMATIC DIAGRAM OF TRANSESTERIFICATION PROCESS | 11 |
| FIGURE 5 – SCHEMATIC DIAGRAM OF LCE FERMENTATION/HYDROLYSIS PROCESS..... | 18 |
| FIGURE 6 – ETHANOL PRODUCTION EFFICIENCY AND YIELD FROM PUBLISHED DATA FOR VARIOUS TIME PERIODS | 21 |
| FIGURE 7 – ETHANOL PRODUCTION CAPACITY AND TOTAL CAPITAL INVESTMENT FOR VARIOUS TIME PERIODS | 22 |
| FIGURE 8 – SCHEMATIC DIAGRAM SHOWING LCE FERMENTATION/TC PROCESS | 29 |
| FIGURE 9 – SCHEMATIC DIAGRAM SHOWING GENERAL PROCESS STEPS FOR FTS | 31 |
| FIGURE 10 – BASIS FOR MID-TERM LCMD - FT TECHNOLOGY CHARACTERIZATION | 34 |
| FIGURE 11 - SCHEMATIC DIAGRAM OF THERMOCHEMICAL ETHANOL PRODUCTION PROCESS | 37 |
| FIGURE 12 – SCHEMATIC DIAGRAM OF THE LCG UPGRADING/PYROLYSIS PROCESS | 43 |
| FIGURE 13 – CAPITAL COST AND PLANT SIZE RELATION FOR FAST PYROLYSIS | 46 |
| FIGURE 14 – LABOR COST SCALING BY FACILITY SIZE FOR PYROLYSIS OIL PRODUCTION | 47 |
| FIGURE 15 – SCHEMATIC DIAGRAM OF RENEWABLE DIESEL PRODUCTION IN PETROLEUM REFINERY | 49 |
| FIGURE 16 – PROJECTED SOYBEAN OIL PRICE | 71 |
| FIGURE 17 – PROJECTED CANOLA OIL PRICE | 71 |

List of Tables

| | |
|--|----|
| TABLE 1 – CURRENT AND ADVANCED BIOFUEL CONVERSION TECHNOLOGIES | 2 |
| TABLE 2 - EXAMPLE COST AND PERFORMANCE OUTPUTS FOR DRY MILL FACILITY | 9 |
| TABLE 3 - EXAMPLE COST AND PERFORMANCE OUTPUTS FOR WET MILL FACILITY | 9 |
| TABLE 4 – BIODIESEL YIELDS FOR VARIOUS FEEDSTOCKS..... | 13 |
| TABLE 5 - EXAMPLE COST AND PERFORMANCE FOR BASE CATALYZED TRANSESTERIFICATION FACILITY | 14 |
| TABLE 6 - EXAMPLE COST AND PERFORMANCE FOR ACID CATALYZED TRANSESTERIFICATION FACILITY | 14 |
| TABLE 7 - TYPICAL BIOCHEMICAL COMPOSITIONS OF VARIOUS FEEDSTOCKS | 17 |
| TABLE 8 - CAPITAL COST BREAK-OUT FOR LCE FERMENTATION/HYDROLYSIS WITH DILUTE ACID PRETREATMENT | 23 |
| TABLE 9 - CONVERSION RATES FOR LCE FERMENTATION/HYDROLYSIS | 24 |
| TABLE 10 - FEEDSTOCK COMPOSITION AND YIELD DATA | 26 |
| TABLE 11 - EXAMPLE COST AND PERFORMANCE - MID TERM DILUTE ACID LCE FERMENTATION/HYDROLYSIS..... | 27 |
| TABLE 12 - EXAMPLE COST AND PERFORMANCE - MID TERM STEAM EXPLOSION LCE FERMENTATION/HYDROLYSIS | 27 |
| TABLE 13 - EXAMPLE COST AND PERFORMANCE FOR FISCHER TROPSCH FACILITY | 36 |
| TABLE 14 - EXAMPLE COST AND PERFORMANCE FOR ETHANOL FROM LCMA SYNTHESIS/TC..... | 40 |
| TABLE 15 - EXAMPLE COST AND PERFORMANCE FOR LCG UPGRADING/PYROLYSIS..... | 47 |
| TABLE 16 - EXAMPLE COST AND PERFORMANCE FOR FAHC - HYDROTREATMENT STAND ALONE PROCESS | 52 |
| TABLE 17 - EXAMPLE COST AND PERFORMANCE FOR FAHC - HYDROTREATMENT CO-PROCESSING..... | 52 |
| TABLE 18 – WATER USE AND YIELD FOR VARIOUS CONVERSION TECHNOLOGIES | 53 |
| TABLE 19 - SUMMARY OF CONVERSION TECHNOLOGY EMISSIONS FROM PUBLISHED LIFE CYCLE ANALYSES..... | 54 |
| TABLE 20 – CO ₂ EMISSIONS (G/GALLON) FROM BIOFUEL CONVERSION BASED ON TECHNOLOGY MODEL DATA | 56 |
| TABLE 21 – OIL CONTENT OF VARIOUS OILSEED FEEDSTOCKS..... | 70 |
| TABLE 22 COMPOSITION OF MSW STREAM COMPONENTS | 72 |

Contributors

Antares Group, Inc.

Ed Gray, President

Christopher Lindsey, Associate Principal

Anneliese Schmidt, Renewable Research Scientist

Acknowledgements

The authors would like to acknowledge the efforts of the following people for their technical review of this report: Jim Kerstetter; Nick Nagel, National Renewable Energy Laboratory; Graham Parker, Pacific Northwest National Laboratory; John Scahill, U.S. Department of Energy Golden Field Office; Richard Hamilton, Ceres, Inc.; Ken Vogel, University of Nebraska; Joe Bouton, The Noble Foundation; Thomas Robb, Abengoa Bioenergy; David Hallberg, PRIME BioSolutions; Bill Carlson, Carlson Small Power Associates.

1 Introduction

This analysis investigates the biofuel conversion technologies that are currently available, as well as technologies currently under development that are far enough along the development path to potentially be available on a commercial basis in the time frame for the strategic assessment (circa 2015). Due to the complexity of the biofuel conversion technologies, there is no simple equation to describe production costs and process yields. As such, spreadsheet models were developed to calculate the costs and yields for selected technologies based on key variables: feedstock type, conversion option and facility size.¹ The technologies were chosen to be representative of the types of biofuel production processes that the assessment team believes can be commercial in the mid term. They are not endorsements of any specific technology and the assessment team believes that the actual plant configurations built will likely incorporate features that could not be foreseen in this study. The technology choices for detailed analysis were based on the relative benefits and challenges for each conversion process, status of the technology, and availability of published data. It is assumed that to have a significant presence in future markets, competing technologies will need to have similar or better yields and similar or lower production costs than the modeled technologies.

Ten biofuel conversion technologies were considered in the assessment. Brief explanations and process descriptions for each of these conversion technologies are given in Table 1. Reference name abbreviations are also included in this table. An overview of the benefits, challenges and status of each technology is provided in Appendix A. Detailed technical descriptions of the current conversion processes are given in Section 2, and descriptions of the advanced technologies are provided in Section 3. A summary of pretreatment requirements for the technologies is given in Appendix B.

From the detailed characterization, the following conversion technologies were selected to include in the supply curve assessment: grain to ethanol production via dry and wet milling, FAME transesterification to produce biodiesel, lignocellulosic ethanol production via enzymatic hydrolysis and fermentation, Fischer Tropsch synthesis, pyrolysis oil production and upgrading, and renewable diesel² production via hydrotreatment of fatty acids. The subsection treating each technology includes a description of the methodology and assumptions used to generate the performance and cost model and an overview of the results.

This report focuses on technologies that convert solid biomass into liquid fuels, as these form the basis for current research and are most likely to be available in the 2015 time period. Never-the-less, it is worthwhile to mention that other conversion technologies and fuels could potentially be incorporated in the transportation market in the future. For example, there is a potential for the use of compressed biogas as an alternative to natural gas for buses or specialized fleets. Biogas can be generated from manure and wastes via anaerobic digestion, and is also a by-product from landfills that can be harnessed for energy. However, such alternatives are not included in this analysis as they are not expected to make up a significant portion of the fuels market in the considered time frame.

¹ Some of the technologies have different process options that can affect the yield and performance of the conversion. For example, renewable diesel can be generated via stand-alone or co-processing conditions, and lignocellulosic ethanol production has several pretreatment process options.

² Renewable Diesel is a hydrocarbon fuel which is generated via hydrotreatment of plant oils and animal fats.

Table 1 – Current and Advanced Biofuel Conversion Technologies

| Reference Name | Feedstock | Conversion Technology | Fuel Type | Size Range* (MGY) | Description |
|---|-------------------------------------|---|------------------------------|-------------------|--|
| Current Representative Technologies | | | | | |
| Grain to Ethanol - Dry Mill | Grains / Starches | Enzymatic Fermentation | Ethanol | 5 to 100 | Dry milling process - grains are ground into a flour, and the starch is converted into sugar with enzymes and fermented to produce ethanol. |
| Grain to Ethanol - Wet Mill | Grains / Starches | Separation and Fermentation | Ethanol | 50 to 300 | Wet milling process - grain separated into components and starch is yeast fermented and distilled. |
| Fatty Acid to Methyl Ester (FAME) | Seed Oil / Waste Oils / Animal Fats | Esterification | Methyl Esters | 1 to 80 | Vegetable oils and fats are filtered and converted via base catalyzed transesterification, producing biodiesel and glycerin which must be separated. |
| Sugar to Ethanol Fermentation | Sugars | Fermentation | Ethanol | 5 to 100 | Sugar crops (like sugar cane) are milled and fermented to produce ethanol. |
| Additional Technologies Projected to be in Use by 2015 to 2025 | | | | | |
| LignoCellulosics to Ethanol - Fermentation/Hydrolysis (LCE Fermentation/Hydrolysis) | Lignocellulosic Biomass | Hydrolysis and Fermentation | Ethanol | 20 to 100+ | Cellulose and hemicellulose converted to sugars via hydrolysis. Various options for hemicellulose conversion (pretreatment). Conversion of sugars to alcohol via fermentation. |
| LignoCellulosics to Ethanol - Fermentation/ThermoChemical (LCE Fermentation/TC) | Lignocellulosic Biomass | Gasification and Fermentation | Ethanol | 50 to 100+ | Gasification to produce syngas, which is then conditioned and compressed. The compressed gas is fermented to ethanol. |
| LignoCellulosics to Middle Distillates - Fischer Tropsch (LCMD - FT) | Lignocellulosic Biomass | Gasification and Fischer Tropsch Synthesis | Middle Distillates, Gasoline | 5 to 100+ | Gasification to produce syngas, which is then cleaned and purified. The clean syngas then undergoes catalytic synthesis (reactor with Co catalyst to maximize diesel fraction). The product is separated and upgraded. |
| LignoCellulosic to Mixed Alcohol - ThermoChemical (LCMA - TC) | Lignocellulosic Biomass | Gasification and Thermochemical Conversion | Mixed Alcohols | 15 to 100+ | Syngas production via biomass gasification, followed by catalytic conversion to mixed alcohols. |
| LignoCellulosics to Gasoline - Upgrading/Pyrolysis (LCG - Upgrading/Pyrolysis) | Lignocellulosic Biomass | Pyrolysis Oil Production and Upgrading via hydrotreatment / hydrocracking | Bio-oil, Diesel, Gasoline | 5 to 100+ | Biomass conversion to bio-oil via fast pyrolysis, co-processing with fossil fuels in petroleum refinery. |
| Fatty Acids to HydroCarbon - Hydrotreatment (FAHC - Hydrotreatment) | Seed Oil / Waste Oils / Animal Fats | Upgrading via hydrotreatment | Renewable Diesel | 5 to 200 | Biomass oils conversion to diesel and other hydrocarbons via hydrotreating methods as in petroleum refinery |

* Size Range gives potential facility sizes expected in the 2015 time frame.

Conversion Technology Model Overview

Conversion technology models were developed to provide an estimate of the cost and performance of the selected technologies in terms of feedstock type and facility size for the GIS based supply analysis for the Western states. These models are based on data available from published studies. An overview of the major assumptions and methodology that apply to the models is given here for reference. Some of the limitations of the models are also discussed.

In general, the key input variables for each model are the feedstock input type and quantity. Key outputs are biofuel yield, capital and O&M costs, and by-product quantity and value. Levelized non-feedstock production costs are also calculated from the key outputs

and economic factors described below. Further analysis will be done by the UC Davis team to generate the biofuel supply curves for the Western Region.

Some of the overall standards for reporting are as follows:

- All reported cost values are given in U.S. 2006\$ unless otherwise noted. Price conversions from other years were performed using the Chemical Engineering Plant Cost Index (CEPCI) and standard relations.
- Models for current commercial grain ethanol and FAME biodiesel production technologies are based on current cost and performance data. As these technologies are already well developed, no projections of future improvements are included.
- Advanced technologies models are based on cost and performance projections available in the literature. The projected costs are typically estimates for the n^{th} plant, with accuracy in the range of +/- 25% to 30%.
- All cost and performance models for the advanced technologies include projections for the 2015 ("mid-term") time period. Models of other time periods were also included for technologies that had additional projections available.³
- Economic factors for calculation of levelized non-feedstock production costs are based on the values used in the CDEAC study, as recommended by the Quantitative Work Group. The production cost uses a Fixed Charge Rate (FCR) of 12.3%, calculated based on an economic lifetime is 25 years and a weighted cost of capital of 9.7%. Additional details on these factors are provided in Appendix C.

Although efforts were made to perform a comparable analysis of each technology it was not always possible to make the same assumptions for each model, since these models are based on data from published studies with varying levels of detail. This can be particularly significant for items which have highly fluctuating market values. For example, electricity costs and fuel prices have increased sharply over the past few years. Where possible, these increased costs have been accounted for by directly substituting current values into calculation of these operational costs. However, in some studies there is not sufficient detail to make this substitution. In these cases the utility costs may be underestimated. In order to make the assumptions for each analysis as transparent as possible, as much detail as practical is included in the report regarding the estimations and calculations for each model.

³ For example, lignocellulosic ethanol production is modeled for 3 time periods – near term (2010), mid term (2015), and long term (2025+).

2 Current Technologies

Ethanol and FAME biodiesel are currently produced in significant quantities using commercial conversion technologies. These technologies include enzymatic fermentation of sugars and starch (grains) to produce ethanol, and transesterification of oils and greases to generate biodiesel.

2.1 Grain to Ethanol Conversion

Ethanol (ethyl alcohol) production is primarily a fermentation process that converts glucose into alcohol.⁴ The main difference between the conversion technologies currently commercially available is the preliminary treatment of the feedstock before fermentation.

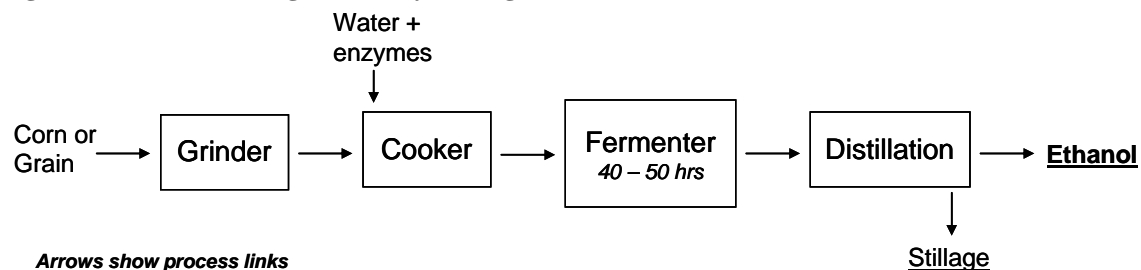
As of February 2007, there were 120 grain ethanol plants in operation in the U.S. with a total production capacity of 5,120 MGY (Industrial Information Resources 2007). More than 98% of the current U.S. ethanol production capacity comes from corn. Of the corn ethanol plants in operation approximately two-thirds are dry-mills. These facilities have high overall ethanol conversion rates and can be built at a scale that is affordable for large agricultural cooperatives and companies focused on ethanol production. About 80% of the U.S. ethanol production comes from dry-mill facilities (Solomon, Barnes and Halvorsen 2007).

2.1.1 Technology Description

Dry Mill Ethanol Facilities

Figure 1 illustrates the basic steps of the dry-milling or enzymatic fermentation process.⁵ In this process, the corn or grain feedstock is initially ground into a flour or fine meal to release the starch. This material is then mixed with water to produce a mash, which is processed in a high temperature cooker with enzymes to convert the starch to sugar and reduce bacterial contamination before fermentation. Ethanol is produced during fermentation, which usually takes 40 to 50 hours (RFA 2005). The ethanol is purified and separated from the stillage during distillation. Further distillation and dehydration (as with molecular sieves) must be done to purify the ethanol for use as a fuel and alter the alcohol concentration.

Figure 1 – Schematic Diagram of Dry Milling Ethanol Production Process



⁴ "Fermentation typically refers to the conversion of sugar to alcohol using yeast under anaerobic conditions. A more general definition of fermentation is the chemical conversion of carbohydrates into alcohols or acids." (Source: Wikipedia)

⁵ A more detailed schematic diagram is given in Appendix D.

The whole stillage (WS) by-product of the dry-milling process, also known as distillers' grain, includes the fiber, oil and protein components of the grain, as well as the non-fermented starch. It is currently sold as an animal feed in a variety of forms. As a result of its high moisture content, stillage is readily perishable and only has a shelf life of 6 to 10 days. To increase durability of the product, thin stillage can be separated and dried to a thick, viscous syrup which is mixed back with the solids to create a feed product known as wet distillers' grains with solubles (WDGS). WDGS, containing 65% moisture, can be used directly as an animal feed product. However, the shelf life of WDGS is only slightly longer than WS (one to two weeks), so end-uses such as feedlots must be nearby (typically within about 50-100 miles of the mill) to avoid spoiling. To increase shelf life and lower transportation costs, WDGS is usually dried to 10-12% moisture to produce a product known as dried distillers' grain with solubles (DDGS). However, the drying process is extremely energy-intensive, and can consume about one-third of the energy requirements of the entire dry mill facility.

DDGS competes with corn, soybean meal, and dicalcium phosphate in animal feed markets as a source of protein and phosphorus. DDGS has a very long shelf life, and is easily transported across the U.S. and all over the world. The U.S. domestic markets account for about 75% of production, while the remaining 25% is produced to satisfy the demand of international markets (Antares Group Inc. 2005a).

As more dry-mill ethanol facilities come on line, the market may become saturated with these types of animal feed products and alternate end-uses or disposal options will need to be explored in order to keep a high value for this by-product. In general, the overall animal feed market is far larger than existing DDGS production capacities, thus theoretically there is room for growth in that area. However, DDGS producers will need to consider expanding into non-traditional feed markets, beyond feed for ruminating animals like dairy beef and cattle. An additional option is increasing exports to other countries, but these can be stifled by high freight costs.

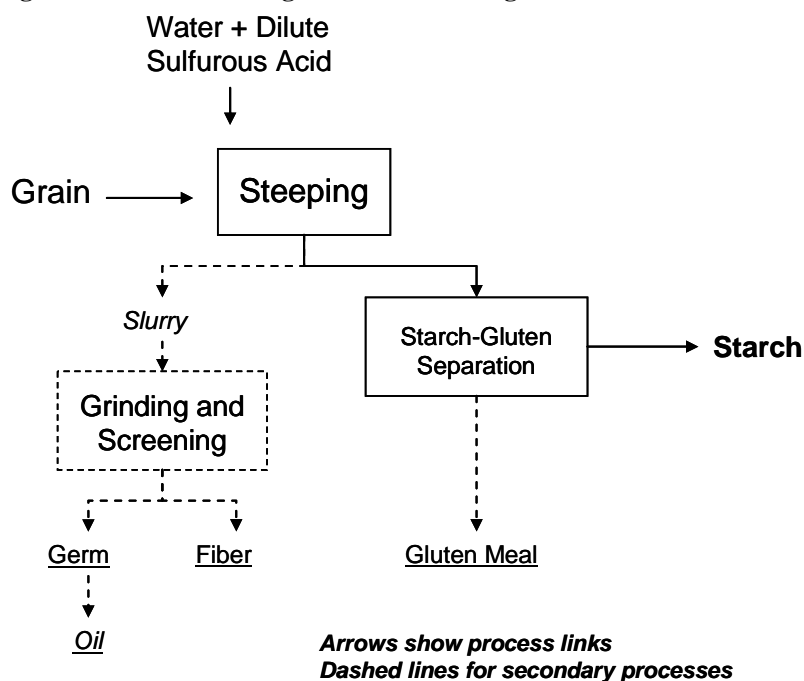
Wet Mill Ethanol Facilities

In the wet milling process, the grain is separated into components (germ, gluten, fiber, and starch) before fermentation, yielding a number of valuable by-products. Figure 2 shows the basic steps of the separation process that is used to produce a fermentable starch.⁶ The grain is first steeped in a mixture of water and dilute sulfurous acid for 24 to 48 hours to facilitate the separation into components (RFA 2005). The germ slurry is then separated, and goes through a process of grinding and screening to separate the fibrous material from the germ. The germ can be sold to crushers, or further processed on-site into an oil product. The gluten is separated from the starch, producing a gluten meal by-product that can be used for animal feed. The separated starch can then be hydrolyzed, fermented and distilled to produce ethanol, using the same steps as those described above for dry milling.

The gluten meal produced from the wet-milling process has high moisture content and is readily perishable. Thus the same problems occur as with the whole stillage by-products of dry milling – the gluten meal must either be used quickly or dried.

⁶ Appendix D has a more detailed schematic diagram of this process.

Figure 2 – Schematic Diagram of Wet-Milling Process for Conversion of Grain to Starch



2.1.2 Performance and cost analysis

Dry Mill Ethanol Facilities

Ethanol yield is directly related to amount of fermentable starch in corn kernels, plant efficiency, plant age, types of equipment, and plant management. Shapouri, Gallagher and Graboski (2002) reported that the average ethanol yield was 2.68 gal/bu for small facilities (<40MGY), and 2.65 gal/bu for large plants. For newer facilities, ethanol yield has increased to 2.8 gallons per bushel.

There have been several technology improvements since the 1970's that have reduced the cost of ethanol production, including a 20% increase in yield of ethanol production from corn from improved biochemistry, higher starch quantities in the kernels, and a 35% reduction in heat energy required for conversion by re-use of process heat (Gallagher, Brubaker and Shapouri 2005). New dry-mill plants use 30,000 Btu of thermal energy and less than 1kWh of electricity to produce 1 gallon of ethanol and by-products (Shapouri and Gallagher 2002). Furthermore, new ethanol plants have minimal discharge of wastewater, and typically do not require wastewater treatment facilities (Shapouri and Gallagher 2002).

There is a clear relationship between capital cost and plant capacity for dry mill facilities, as investigated in Gallagher, Brubaker and Shapouri (2005). This study analyzed several dry mill facilities of varying sizes constructed over a 25 year period, and found that capital costs decrease on a \$/gallon basis with increasing size up to an optimum capacity of about 55-74 MGY, beyond which costs increase with size. The minimum capital cost is \$1.14 per gallon ethanol production capacity (Gallagher et al. 2005).

In 2004, there were about 40 large dry-mill ethanol plants (40-100 MGY) in operation or under construction (Shapouri and Gallagher 2005). The capital costs for new plants at this time ranged from \$1.05 to \$3.00 per gallon of capacity. As expected, the cost for expansion

of existing plants was much lower, from \$0.20 to \$1.00 per gallon capacity (with an average of \$0.50/gal).

According to Shapouri, Gallagher and Graboski (2002) DDGS made up 70% of all distiller's grain products from dry mill facilities, while WDGS and MDGS (modified distillers grains) accounted for 21% and 9%, respectively (all on a dry basis). Furthermore, 17 of 21 ethanol facilities surveyed indicated that they sold wet and modified distiller grains.

Wet Mill Ethanol Facilities

Since wet mills are more complex than dry mills, they have a different plant-size to capital cost relationship (Gallagher, Brubaker and Shapouri 2005). Wet mills also have higher capital and O&M costs. However, energy expenses tend to be lower in wet mills than dry mills as they typically include cogeneration of steam and electricity (Shapouri, Gallagher and Graboski 2002).

Wet mills have a number of valuable by-products, including corn gluten feed, corn gluten meal, and corn oil. The separated starch component can either be fermented to produce ethanol, or it can be further processed into corn syrup and sugar. In this analysis we assume that all of the starch is fermented in an ethanol producing wet mill facility. However, it is important to note that the flexibility of product distribution from wet mills can insulate these facilities from large price fluctuations for one of more of the potential products. For a value of \$2/gallon or higher, ethanol production from a wet mill is competitive with corn syrup production.

Outline for model analyses

The key values and assumptions used to calculate the current technology analysis and economic performance model for wet and dry mills are described below. An example of the detailed analysis is shown in Appendix G.

- Feedstock input quantity is a variable, and should be given on an as-received basis (typically about 15% moisture). Dry mills can use corn or sorghum, while wet mills use only corn.
- Applicable facility size ranges are 5 MGY to 100 MGY for dry mills, and 50 MGY to 300 MGY for wet mills, based on data from Gallagher et al. 2005.
- Ethanol yield is calculated based on the average of values reported in the literature. New dry mills typically produce 2.8 gallons per bushel of corn (100 gal/ton), and wet mills produce about 2.5 gal/bu (89.3 gal/ton) (Shapouri and Gallagher 2005; McAloon et al. 2000, Butzen and Hobbs 2002).
- The conversion efficiency of corn to fuel is estimated to be 64% (on an energy basis) for a dry mill and 57% for a wet mill. This is calculated using a typical HHV for ethanol of 84,000 Btu/gal, and a HHV for corn of 6,600 Btu/lb (at 15% moisture content).
- It is assumed that all distiller's grains from dry mill facilities are converted to DDGS. It is estimated that dry mill facilities produce 6.7 lb DDG per gallon of ethanol produced (McAloon et al. 2000).
- The wet mills by-products include corn gluten feed, corn gluten meal, and corn oil, of which 11.4 lb, 3 lb, and 1.6 lb are produced from 1 bushel of corn, respectively (Butzen and Hobbs 2002).⁷
- The amount of CO₂ production from fermentation for a dry mill facility is estimated to be 17.5 lb per bushel of corn (Antares Group Inc. 2005a; McAloon et al. 2000). The CO₂ production for wet mills is based on the same value, but is scaled by the ethanol conversion rate.

⁷ 1 bushel of corn is equivalent to 56 lbs.

- The average water requirement for a dry mill facility is 4.7 gallons of water per gallon of ethanol produced (Shapouri & Gallagher 2005). Based on data from Shapouri, Gallagher and Graboski (2002), wet mills use 5.2 times as much water as dry mills.
- Capital cost data for dry mill facilities is based on the relationship derived from Shapouri and Gallagher (2005). The minimum capital cost on a \$/gal basis occurs at 65 MGY. For larger facilities the cost increases with capacity.
- Wet mill capital costs are based on data from Whims 2002, using a scaling factor of 0.6.⁸ The capital for a wet mill facility is higher than for a dry mill facility due to additional process equipment requirements.
- Average wholesale values of the co-products are based on current data from United States Department of Agriculture (USDA 2007). The value of DDGS is \$90 - \$115 per ton (similar to data from McAloon et al. 2000 and Shapouri and Gallagher 2005). Corn gluten feed is \$40-60/ton, corn gluten meal is \$335-350/ton, and corn oil is \$0.31-0.32/lb. These current wet mill co-product values from USDA are comparable to those reported for 2003 in Shapouri and Gallagher (2005).
- Consumables for both processes include enzymes, yeast, chemicals, and denaturant. The consumables cost for small dry mills (< 40 MGY) is \$0.13/gal, while for large dry mills (40 to 100 MGY) the cost is \$0.12/gal (Shapouri and Gallagher 2005). The cost of consumables for wet mills is \$0.19/gal (Shapouri, Gallagher and Graboski 2002).
- Utilities include electricity, fuels, water and waste management. The cost for small dry mills is \$0.27/gal and for large dry mills the cost is \$0.19/gal (Shapouri and Gallagher 2005). Average electricity use is 1.19 kWh/gallon of ethanol produced, and the average heat use is 34,800 Btu per gallon of ethanol. The cost of electricity for the dry mill facilities has been updated to 2005 average wholesale value for the industrial sector of 5.7¢/kWh. New dry mill facilities have minimal wastewater discharge. Wet mill utility costs are \$0.19/gal (Shapouri et al. 2002).⁹ Wet mill energy usage is lower than dry mills as they typically employ cogeneration of steam and electricity.
- Annual labor costs for small dry mill facilities are \$0.077/gal, and \$0.060/gal for large facilities (Shapouri and Gallagher 2005). Labor costs for wet mills are \$0.096/gal, based on data from Shapouri, Gallagher and Graboski 2002. No scaling included for the wet mill costs as the data represents various size facilities.
- Other fixed O&M costs include maintenance, administrative costs, and other costs. For dry mills, other fixed costs are equivalent to \$0.090/gal for small facilities and \$0.106/gal for large facilities (Shapouri and Gallagher 2005). For wet mills other fixed costs is \$0.108/gal (Shapouri, Gallagher and Graboski 2002).
- The levelized non-feedstock production cost of ethanol is calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results for current ethanol production processes are shown in Table 2 and Table 3. These tables show results for the range of applicable facility sizes based on selected input quantities, and illustrate the high and low end costs. In general, the analysis shows that ethanol production from wet mill facilities has lower costs than the dry mills as a result of the co-product value. However, it is important to note that the value for these co-products will fluctuate as the market changes. For similar sized facilities the wet mills have higher capital and O&M costs than dry mills, as the process is

⁸ This scaling factor was derived from the Whims (2002) data.

⁹ Electricity prices have not been updated to reflect current market value as there was not sufficient detail in the Shapouri et al. (2002) to perform the calculations.

more complex. It should also be noted that both tables show results only for the non-feedstock portion of the production costs.

Table 2 - Example Cost and Performance Outputs for Dry Mill Facility

| Sample Model Results - Grain Ethanol (Dry Mill) | | |
|--|---------------|---------------|
| Feedstock Input (as rec'd ton/yr) | 50,000 | 1,000,000 |
| Ethanol Yield (MGY) | 5 | 100 |
| DDG Yield (ton/yr) | 16,750 | 335,000 |
| CO2 Stream (ton/yr) | 15,625 | 312,500 |
| Water Consumption (1000 gal/yr) | 23,500 | 470,000 |
| Capital Cost (Million \$) | \$12.6 | \$167.2 |
| Annual O&M Costs (Million \$/yr) | \$2.9 | \$49.9 |
| By-Product Credit (Million \$/yr) | \$1.7 | \$34.3 |
| Non-feedstock Production Cost (\$/gal) | \$0.55 | \$0.36 |

Table 3 - Example Cost and Performance Outputs for Wet Mill Facility

| Sample Model Results - Grain Ethanol (Wet Mill) | | |
|--|---------------|---------------|
| Feedstock Input (as rec'd ton/yr) | 560,000 | 3,360,000 |
| Ethanol Yield (MGY) | 50 | 300 |
| Corn Gluten Feed Yield (ton/yr) | 114,000 | 684,000 |
| Corn Gluten Meal Yield (ton/yr) | 30,000 | 180,000 |
| Corn Oil Yield (ton/yr) | 16,000 | 96,000 |
| CO2 Stream (ton/yr) | 156,250 | 937,500 |
| Water Consumption (1000 gal/yr) | 1,222,000 | 7,332,000 |
| Capital Cost (Million \$) | \$139.1 | \$407.7 |
| Annual O&M Costs (Million \$/yr) | \$29.5 | \$176.9 |
| By-Product Credit (Million \$/yr) | \$26.1 | \$156.3 |
| Non-feedstock Production Cost (\$/gal) | \$0.41 | \$0.24 |

According to USDA AMS 2007, the average price for corn in late 2007 was around \$135/ton (including an estimated delivery cost of \$10/ton).¹⁰ This implies a feedstock cost for ethanol production of \$1.35/gallon for dry mills and \$1.51/gallon for wet mills, based on the conversion yields used in the model. The resulting ethanol production costs are about \$1.71 to \$1.92 per gallon. For comparison, current FOB ethanol prices are around \$1.75-\$2.15/gal (USDA AMS 2007). This is very similar to the results of the ethanol production models reported here.

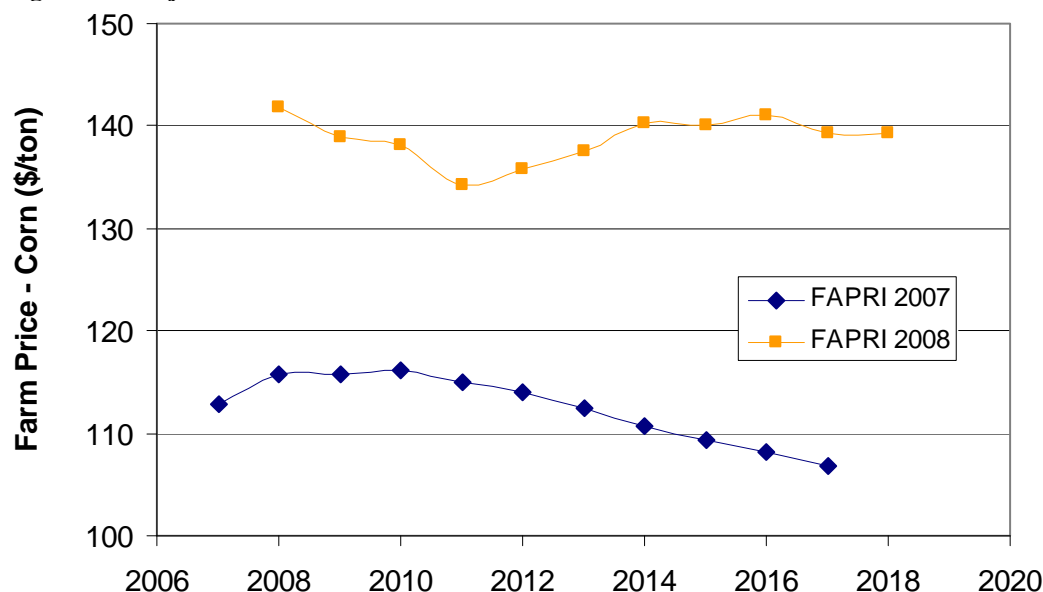
Projected prices for corn at the farm gate from FAPRI's 2007 and 2008 U.S. and World Agricultural Outlook are shown in Figure 3. Since the 2008 data was not available at the time of analysis, the 2007 data is used in the report. Note that although the more recent data shows higher corn price projections, both data sets have relatively flat price escalation curves.¹¹ The higher corn prices from the 2008 data likely reflect increasing costs of energy. As such, using the 2007 corn price data for analysis is consistent with the energy price projections employed in this analysis.

According to the 2007 data, corn is project to cost about \$119/ton in 2015 including the same \$10/ton transportation cost as above. This leads to ethanol production costs of \$1.55 to \$1.74 per gallon for dry mills and \$1.57 to \$1.74 per gallon for wet mills.

¹⁰ Corn price based on data from USDA AMS 2007. Estimated delivery cost based on Antares experience, using a typical cost of \$0.20/ton-mile for truck transportation and a distance of 50 miles.

¹¹ The projected corn price escalation from 2008 to 2015 is -5.6% with the 2007 FAPRI data, and -1.3% with the 2008 FAPRI data.

Figure 3 – Projected Corn Prices



Data from FAPRI 2007 and FAPRI 2008

2.2 Fatty Acid to Methyl Esters: Transesterification (Biodiesel)

There are three basic conversion routes for fatty acid methyl ester (FAME) production: base catalyzed transesterification of oil with alcohol; direct acid catalyzed transesterification of oil with methanol; or conversion of oil to fatty acids, followed by acid catalyzed transesterification to alkyl esters (EUBIA 2006; NBB n.d.). The first option tends to be the most economic for virgin feedstocks and as such is most commonly used to produce esters on a commercial scale.¹² There are several reasons that favorably affect the process economics, including low temperature and pressure processing requirement, high conversion rates (around 98%), lower reaction times, and direct conversion to methyl esters without producing any intermediate compounds (EUBIA 2006).

Acid catalyzed transesterification is expected to be the preferred method for conversion of waste oils, since it is less sensitive to free fatty acids in the feedstock. This conversion method seems to be more economical than base catalyzed transesterification of waste oils, which requires an extra pretreatment step to remove impurities (Zhang et al. 2003b). Although acid catalyzed transesterification has not yet been optimized for commercial scale production, it is expected that this reaction method could be available by 2015.

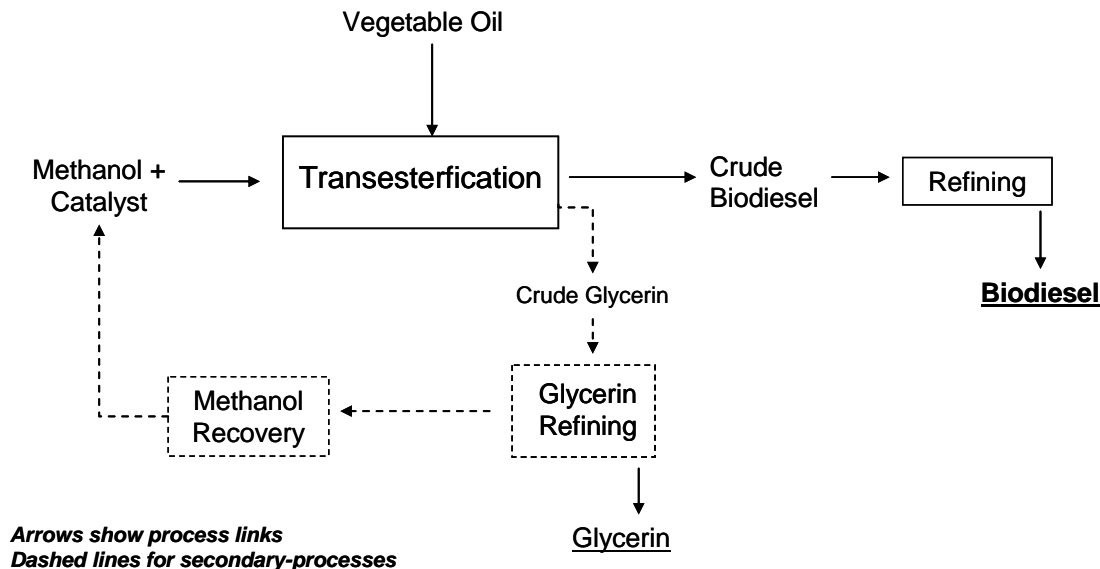
2.2.1 Technology Description

Figure 4 illustrates the main process steps for transesterification of vegetable oil. Before transesterification, the vegetable oils and fats are filtered to remove water and contaminants. The oil is then mixed with an alcohol (often methanol, because of its low cost) and a catalyst (usually sodium hydroxide or potassium hydroxide for base catalyzed reaction, and sulfuric acid for the acid catalyzed process) in a simple closed reactor system at low temperature and pressure. For base catalyzed process the molar ratio of methanol to oil is about 6:1, while for an acid catalyzed process the ratio is about 50:1 (Zhang et al.

¹² Virgin feedstocks include vegetable (seed) oils and animal fats.

2003a). The mixture is left to settle in the reaction vessel for 1 to 8 hours, after which the glycerin and crude biodiesel (methyl esters) are separated. The majority of the alcohol is recovered from the glycerin by-product and recycled back into the system. The biodiesel is purified by a washing process to remove residual catalyst and soaps.

Figure 4 – Schematic Diagram of Transesterification Process



The purity and yield of the methyl esters are affected by the molar ratio of glycerides to alcohol, type of catalyst, reaction time, reaction temperature, and the free fatty acids and water present (Vicente, Martinez and Aracil 2007; Ma and Hanna 1999).¹³ The presence of water causes soap formation during the reaction, and free fatty acids significantly reduce ester yields (particularly for base catalyzed reaction). The formation of soap is an undesirable by-product as it reduces the biodiesel yield and makes glycerol separation more difficult (Vicente, Martinez and Aracil 2007).

Experiments involving the transesterification of sunflower oil feedstock showed that the initial catalyst concentration was the most important factor for biodiesel purity and yield. The catalyst concentration was found to positively affect the biodiesel purity, and negatively affect the yield (Vicente, Martinez and Aracil 2007). In other words, a higher initial catalyst concentration produced more pure biodiesel in smaller quantities than a lower catalyst concentration. Vicente, Martinez and Aracil (2007) also found that higher temperatures negatively affected biodiesel yield and positively affected the purity.

Although under similar conditions the acid catalyzed transesterification process has a slower reaction rate than the base catalyzed process, this can be overcome by a number of methods including longer reaction times, using a larger quantity or increased concentration of acid catalyst, or a higher reaction temperature (Zhang et al. 2003a). However, the acid catalyzed process has not yet been done on a commercial scale.

¹³ Fatty acids can be bonded to another molecule (such as glycerol in triglycerides). Free fatty acids are not attached to another molecule, and are a type of impurity in oils used for biodiesel production. Ma and Hanna (1999) note that if oils are not dry and free fatty acids (<0.5%), ester yields may be significantly reduced.

2.2.2 Performance and cost analysis

A study by Haas et al. (2006) investigated the economics of a 10 MGY acid catalyzed FAME facility using virgin oil feedstock. They found that the largest portion of the capital cost was storage tanks to hold a 25 day supply of oil feedstock, which made up about 1/3 of equipment costs. An analysis by Zhang et al. (2003b) showed that the transesterification reactor vessels and distillation columns were the largest portions of the equipment costs, particularly for the acid catalyzed process (which has a much larger methanol requirement and requires 2 reactors and stainless steel vessels). However, their analysis did not include the cost of feedstock storage vessels.

For a transesterification processes using virgin oil, the feedstock is the bulk of the production costs. For example, in the analysis by Haas et al. (2006), soybean oil made up 88% of the calculated production costs. Furthermore, production cost was found to vary linearly with feedstock costs, and inversely with market value of glycerol. In their analysis, Zhang et al. (2003a) showed that the base catalyzed process with virgin oil was the most simple and had the least process equipment, but also had very high feedstock costs. Zhang et al. (2003b) showed that the economics of acid catalyzed process using waste oil feedstocks can be competitive with base catalyzed process using virgin feedstocks. Although the acid catalyzed process has higher capital and O&M costs, the feedstock is much less expensive. In their analysis, Zhang et al. (2003b) showed that plant capacity and feedstock oil price had the largest effect on production costs of biodiesel for both types of processes.

Outline for model analysis

The key values and assumptions used to calculate the current technology analysis and economic performance model for transesterification are described below. An example of the detailed analysis is shown in Appendix G.

- Feedstock type and input quantity are variables. The feedstock options are virgin oil, animal fats, and yellow grease, and the applicable size range is 1 to 80 MGY.
- Biodiesel yield is calculated based on the average of values reported in the literature for each feedstock type (summarized in Table 4). In general, the virgin feedstocks have higher yields than yellow grease. Reported yields from virgin oil are from 246.2 to 270 gal/ton (90% - 99% conversion on a mass basis), and for yellow grease the yields are 234.9 to 263.2 gal/ton (86% - 97% conversion).¹⁴
- The conversion efficiency of oil/fat to fuel on an energy basis is estimated to be 88-94%, depending on the feedstock. This is calculated using an average HHV of oil feedstocks of 16,500 Btu/lb, and a typical biodiesel heating value of 117,000 Btu/gal.
- Glycerin is the only significant by-product of transesterification. The average production rate is 0.8 lb glycerin per gallon of biodiesel produced from virgin feedstocks (Bender 1999, Haas et al. 2006). These studies refer to the production of a partly purified glycerol product, which contains about 80% glycerol by mass. Although waste oils may have slightly higher glycerin yields, this has not been included in the analysis. The value of the glycerol is estimated to be \$0.05/lb, based on current market price data from Nilles (Sept. 2006) and personal communication with Leland Tong (National Biodiesel Board contributor). Note that this is a particularly variable market and it is difficult to predict what the product value will be in the future.
- Annual water consumption is estimated to be 0.25 lb per gallon biodiesel produced, based on data from Haas et al 2006.
- Capital cost for base catalyzed transesterification is based on data for a 10 MGY facility from Haas et al. 2006. The acid catalyzed facility cost is based on equipment cost from Zhang et al. 2003b for a 2.4 MGY facility, with added cost for storage facilities from Haas

¹⁴ Sources: Haas et al. 2006, Sheehan et al. 1998, Zhang et al. 2003a, Canakci and van Gerpen 2001.

et al. 2006. Additionally, the multiplier from Haas et al. (2006) was used to calculate the total capital investment for this facility. Capital cost is scaled using a scaling factor of 0.6, based on data from Bender (1999).

- In general, the capital costs of acid catalyzed transesterification are estimated to be higher than base catalyzed transesterification because the process uses a greater quantity of methanol, which requires more and larger transesterification reactors and a larger distillation column (Zhang et al. 2003b). Additionally, the acid catalyzed process is much more corrosive and requires expensive stainless steel process equipment. The increased methanol requirement and feedstock pretreatment also leads to increased O&M costs for the acid catalyzed process.
- Labor costs are calculated based on data for a virgin oil facility from Haas et al. 2006, but the operator salary was increased to \$20/hr from \$12.5/hr based on Antares experience. Values in Zhang et al. (2003b) suggest that acid catalyzed facilities have higher labor requirements, so Antares estimated there would be one additional operator per shift for a 10 MGY facility.¹⁵ The total labor cost is scaled by facility size, based on scaling data available from pyrolysis oil production facilities.
- Annual non-labor fixed O&M costs are calculated as a percentage (2.1%) of the capital cost, following Haas et al. (2006).
- Annual Variable O&M costs include consumables (methanol, sodium methoxide, hydrochloric acid, sodium hydroxide, and water) and utilities (natural gas, WWT, waste disposal, and electricity). The annual cost for utilities are calculated on a \$/gallon basis of biodiesel produced, based on data from Haas et al. 2006 for the base catalyzed process and Zhang et al. 2003b for the acid catalyzed process.¹⁶ The annual cost of consumables for both processes is based on data from Haas et al. (2006). However, the quantity and price of the methanol and sulfuric acid for the acid catalyzed process are substituted using data from Zhang et al. (2003b).
- The levelized non-feedstock production costs for biodiesel are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Table 4 – Biodiesel yields for various feedstocks

| Feedstock (1) | Yield range (gal/ton) | Average Yield (gal/ton) | Sources |
|-------------------|-----------------------|-------------------------|--|
| Virgin Oil (2) | 246.4-270.0 | 258.2 | Haas et. al 2006, Sheehan et al. 1998, Zhang et al. 2003a, Canakci and van Gerpen 2001 |
| Animal Fats | 266.3 | 266.3 | Bender 1999 |
| Yellow Grease (3) | 234.9 - 263.2 | 249.1 | Zhang et al. 2003a, Canakci and van Gerpen 2001 |

1) All data based on feedstocks with negligible water content.

2) Based on data for degummed soybean oil. High end of the yield range from Haas et al 2006, which assumes a negligible free fatty acid content in the feedstock.

3) Conversion yield for waste greases based on acid catalyzed reaction. For comparison, the transesterification reaction for virgin feedstocks use base catalysts.

Model Results:

Some of the key cost and performance results for current FAME production processes are shown in Table 5 and Table 6. These tables show results for the range of applicable facility sizes based on selected input quantities, and illustrate the high and low end costs. In

¹⁵ For comparison the base catalyzed facility had 2 operators per shift.

¹⁶ Electricity prices have not been updated to reflect current market value as there was not sufficient detail in the published studies to perform the calculations.

general, the analysis shows that the levelized non-feedstock production cost of biodiesel for the base catalyzed transesterification process is significantly lower than the acid catalyzed process, as expected. However, the feedstock cost will be much more expensive for the virgin oil transesterification facility (typically waste grease is about half as expensive as virgin oil).

Table 5 - Example Cost and Performance for Base Catalyzed Transesterification Facility

| Sample Model Results - FAME Biodiesel w/ Virgin Oil | | |
|--|---------------|---------------|
| Feedstock Input (ton/yr) | 5,000 | 310,000 |
| Biodiesel Yield (MGY) | 1.3 | 80.0 |
| Crude Glycerin Yield (ton/yr) | 516 | 32,020 |
| Water Consumption (1000 gal/yr) | 39 | 2,400 |
| Capital Cost (Million \$) | \$3.3 | \$39.5 |
| Annual O&M Costs (Million \$/yr) | \$0.5 | \$20.1 |
| By-Product Credit (Million \$/yr) | \$0.05 | \$3.2 |
| Non-feedstock Production Cost (\$/gal) | \$0.67 | \$0.27 |

Table 6 - Example Cost and Performance for Acid Catalyzed Transesterification Facility

| Sample Model Results - FAME Biodiesel w/ Waste Grease | | |
|--|---------------|---------------|
| Feedstock Input (ton/yr) | 5,000 | 310,000 |
| Biodiesel Yield (MGY) | 1.2 | 77 |
| Crude Glycerin Yield (ton/yr) | 498 | 30,880 |
| Water Consumption (1000 gal/yr) | 37 | 2,310 |
| Capital Cost (Million \$) | \$5.2 | \$61.5 |
| Annual O&M Costs (Million \$/yr) | \$1.0 | \$45.8 |
| By-Product Credit (Million \$/yr) | \$0.05 | \$3.1 |
| Non-feedstock Production Cost (\$/gal) | \$1.26 | \$0.65 |

According to USDA AMS (2007), the current cost for crude soybean oil is about 32.5 cents per pound (\$2.5/gallon), and the cost for biodiesel is around \$3.10/gallon. This implies a non-feedstock production cost of \$0.60, similar to the projections from the model for small base catalyzed transesterification facilities. Haas et al. (2006) estimate a biodiesel production cost of \$2.00/gallon for a 10 MGY facility using crude soybean oil at \$1.81 per gallon (23.5¢/lb). For a similar size facility using the same feedstock price, the FAME biodiesel model gives a production cost of \$2.20/gal.

The Food and Agricultural Policy Research Institute (FAPRI) 2007 U.S. and World Agricultural Outlook, projects 2015 soybean oil prices will be about \$681.40/ton (34.1¢/lb). Including \$10/ton for delivery, this gives production costs of \$2.98 - \$3.36 per gallon for biodiesel.¹⁷ Note that this does not include the cost of distribution, which would be included in the retail fuel price. Refer to Appendix E for details on the conversion of oilseed crops to oil.

2.3 Sugar to Ethanol Fermentation

The conversion of sugars (like sugarcane) to ethanol is simpler than the conversion of starches as it requires fewer process steps. This process is very similar to the dry milling process described above except that no cooking is required. The feedstock is milled and

¹⁷ Delivery cost estimated for truck transportation, using \$0.20/ton-mile and a 50 mile delivery radius. Including delivery, the soybean oil price is \$2.66/gallon.

pressed, and then fermented to generate ethanol. The resulting product is purified during distillation, and can be dehydrated to change the alcohol concentration.

Ethanol production directly from sugars is generally only performed in locations where there is a large quantity of available feedstocks, such as sugarcane crop in Brazil and sugar beets in parts of Europe. According to the USDA National Agriculture Statistics Service (NASS), the entire U.S. produced about 34 million tons of sugar beets and 30 million tons of sugarcane in 2006 (USDA NASS 2007).¹⁸ For comparison, the U.S. soybean crop from 2006 was nearly 90 million tons and the corn crop was 300 million tons (USDA NASS 2007). As a result of limited feedstock availability, sugar crops are not a significant source of ethanol production in the U.S.

¹⁸ The top states with sugarcane crops are Florida, Louisiana, Hawaii and Texas, while the top sugar beet growing states are Minnesota, North Dakota, Idaho, Michigan and California (USDA NASS 2007).

3 Technologies Ready for Commercialization in the Midterm (2010 – 2025)

There are a number of advanced biofuel technologies being developed that may be available in the 2010 to 2025 time frame. One of the key differences between these and the current commercial biofuel technologies is the types of biomass feedstocks used for production. While the current technologies typically use grain and oil seed crops (which are expensive and compete with food usage), many of the advanced technologies use lignocellulosic biomass such as wood, switchgrass and agricultural residues. Although these biomass feedstocks are generally more difficult to convert to biofuels, they are not a human food source and can be much less expensive than grain and seed crops.

Cost and performance models for several of the advanced technologies have been developed. Since the advanced technologies do not currently have commercial plants in operation, these models are based on cost and performance projections available in the literature. The projected costs are typically estimates for the n^{th} plant, with accuracy in the range of +/- 25% to 30%. All cost and performance models for the advanced technologies include projections for the 2015 ("mid-term") time period. Models of other time periods were also included for technologies that had additional projections available in published studies.

3.1 Lignocellulosics to Ethanol: Hydrolysis and Fermentation

Lignocellulosic biomass consists mainly of cellulose, hemicellulose and lignin. This includes woody biomass, grasses, and agricultural residues. Lignocellulosic biomass typically contains 40-60 wt% cellulose (dry weight) and 20-40 wt% hemicellulose, both which can be hydrolyzed to sugars and fermented to produce ethanol. Most of the remaining fraction (10-25 wt%) is lignin, a complex polymer which is resistant to biological degradation and cannot be fermented, but can be thermochemically treated or used directly to produce heat and/or power.¹⁹ There are also small amounts of proteins, lipids, and ash.

The chemical composition of the biomass (i.e. the lignin/hemicellulose/cellulose ratio) is a major factor in the ethanol yield. Materials with high sugar content (cellulose and hemicellulose) have the potential to produce the greatest amount of ethanol. Table 7 shows typical biochemical compositions for various biomass feedstocks. Most woody biomass has about 27% lignin, while grasses such as switchgrass have about 18% (Hamelinck, van Hooijdonk and Faaij 2005). Switchgrass has a large fraction of ash and other materials which can not be converted to ethanol or used for heat and power production.

¹⁹ The higher heating value (HHV) of lignin is 10,495 +/- 515 Btu/lb (dry), whereas the (hemi)cellulose and other components have a much lower HHV around 7,310 Btu/lb (dry) (Hamelinck, van Hooijdonk and Faaij 2005).

Table 7 - Typical biochemical compositions of various feedstocks²⁰

| Feedstock Composition (wt%) | | | | |
|-----------------------------|---------------|-----------|--------|--------|
| Feedstock | Hemicellulose | Cellulose | Lignin | Other |
| Agricultural Resources | 23-26% | 33-40% | 17-24% | 10-27% |
| Switchgrass | 25% | 32% | 18% | 25% |
| Hardwoods (i.e. Poplar) | 13-19% | 42-50% | 26-28% | 3-19% |
| Pine (Softwood) | 22% | 45% | 28% | 6% |

Sources: U.S. DOE EERE Biomass Program Jan 25, 2006; Hamelinck, van Hooijdonk and Faaij 2005

3.1.1 Technology Description

Production of ethanol from lignocellulosic materials is similar to production from grains, except that it is more difficult to break down the feedstock into fermentable sugars. In other words, hydrolysis of lignocellulose is more difficult than hydrolysis of starch.²¹ The hydrolysis or saccharification of cellulose produces glucose (a sugar containing 6 carbon atoms), and hydrolysis of hemicellulose produces various 5 carbon and 6 carbon sugars.

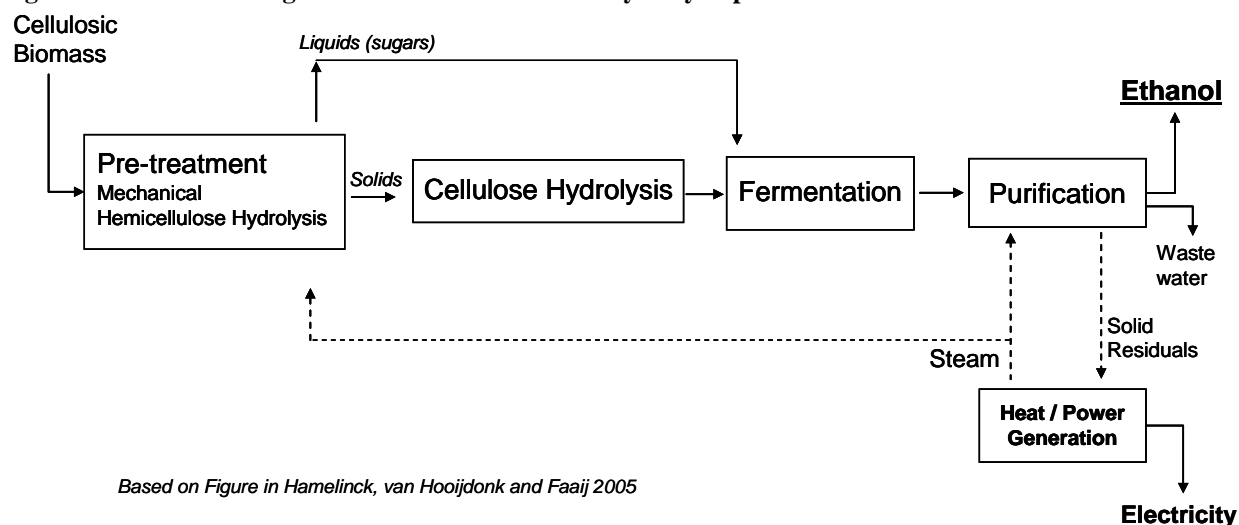
Figure 5 shows the basic process steps for the conversion of LignoCellulosics to Ethanol via Hydrolysis and Fermentation (LCE Fermentation/Hydrolysis). The biomass is first pretreated to clean and size the materials and alter the cell structure so that it is more accessible to conversion. Hemicellulose hydrolysis is part of the pretreatment, and can be performed by a variety of methods (see below). After pretreatment, the solids (cellulose and lignin) are filtered and pressed and undergo cellulose hydrolysis, while the sugar-containing liquids are diverted to fermentation. Cellulose hydrolysis is the main hydrolysis step, in which the cellulose is converted to sugars. The primary methods for this reaction are described below. After hydrolysis, the sugars are converted to ethanol via fermentation, and the ethanol is recovered in a distillation column and concentrated. The lignin and other solid residuals can be thermochemically treated (gasified) or used directly to produce heat/power via combustion. Unfermented sugars in liquid effluent can also be dried and fired in a boiler. An alternate use of the lignin is to produce high-value co-products such as high-octane hydrocarbon fuel additives, but these conversions are complex and expensive. Note that although each step is shown below as a separate process, integration of two or more process steps is a potential way to reduce costs and increase overall efficiencies. In the descriptions below each major process step is discussed separately in order to clarify the difference aspects of the process.

It is important to mention another potential method of ethanol production from lignocellulosic materials; hemicellulose conversion of biomass in a forest product biorefinery. In this process sugars are extracted from the hemicellulose component of the biomass and fermented to ethanol. The rest of the biomass is then used to produce high value pulp and paper products. Production of lignocellulosic ethanol as a by-product of the forest products industry is very appealing, as the required biomass infrastructure is already well developed. Although there is significant interest in this type of biorefinery, there was not sufficient data publicly available for a detailed characterization of this process for this analysis. However, the inclusion of this process will be re-evaluated for future work, as additional technical information is published.

²⁰ "Other" includes ash, acids, and extractives (low molecular weight organic materials like aromatics and alcohols).

²¹ Hydrolysis is the breakdown of organic materials through the use of water.

Figure 5 – Schematic diagram of LCE Fermentation/Hydrolysis process



Pretreatment and Conditioning

Pretreatment is an important part of the conversion process because it has a very significant impact on the ethanol yield. It is also a very energy intensive step – sizing the biomass can take up to one third of the total energy requirement for the process. The hemicellulose saccharification is also included in pretreatment step, as it helps to make the cellulose easier to convert to sugars. In fact, cellulose hydrolysis without the hemicellulose hydrolysis pretreatment step typically has yields below 20%, while after this pretreatment step the cellulose hydrolysis can have yields greater than 90%. Some of the primary hemicellulose hydrolysis methods are described below.

- **Chemical pretreatment processes such as dilute acid:** The acid catalyzed hydrolysis process uses dilute sulfuric, hydrochloric, or nitric acid.²² This method is currently available and gives reasonably high sugar yields (75-90%). However, the acid needs to be removed and neutralized before fermentation which creates a lot of gypsum for disposal. This method also requires very small biomass particle sizes, typically about 1 to 3mm, as smaller particles have larger percentages of surface area available (Hamelinck, van Hooijdonk and Faaij 2005).

Acids need to be removed before enzymatic hydrolysis and fermentation. This is done by separating remaining solids from the liquid, followed by washing and pressing. Acids are then removed or recovered from the liquid fraction via membrane separation (about 80% effective recovery) or by continuous ion exchange (97% recovery with 2% of sugars lost). Low acid concentrations can be neutralized by adding lime, causing a conversion to gypsum which can then be filtered out for disposal.²³ Conditioning is also used to remove inhibitors produced during pretreatment, as they reduce ethanol yields.

- **Physical pretreatment processes such as steam explosion and liquid hot water (LHW):** In steam explosion biomass is heated using high pressure steam for several minutes, and the reaction is stopped by sudden decompression to atmospheric pressure. This process results in sugar yields around 45-65% (Hamelinck, van Hooijdonk and Faaij

²² Note that there is also a concentrated acid hydrolysis process, but it is currently very expensive and will not be discussed in detail here.

²³ There is typically 0.02 kg gypsum produced per kg feedstock after recovery. With acid catalyzed hydrolysis with no acid recovery this can be as high as 0.6 - 0.9 kg of gypsum per kg of feedstock.

2005). LHW uses compressed hot liquid water to hydrolyze the hemicellulose, resulting in high yields of 88-98%. These high yields and the fact that no acid or chemical catalyst required make the process extremely attractive. However, LHW is currently in the demonstration stage of development, and is not expected to be available for large scale commercial projects for another 5 years. Biomass feedstock size required for these processes are around 19mm, much larger than the dilute acid hydrolysis step (Hamelinck, van Hooijdonk and Faaij 2005). This significantly lessens the pretreatment energy requirement.

- **Biological pretreatments (Fungi):** The process is characterized by low energy use and mild environmental conditions. However, it also has low conversion yields and long reaction times. This method tends to be more effective in combination with chemical treatments (Hamelinck, van Hooijdonk and Faaij 2005).

Cellulose Hydrolysis

Cellulose hydrolysis is typically performed via **enzymatic hydrolysis**. In this process, cellulase enzymes convert the cellulose to sugars under mild process conditions.²⁴ This method results in relatively high yields (75-85%) and has lower maintenance costs than acid hydrolysis (described below). Cellulase is currently a very expensive part of ethanol production, although there are research activities that focus on lower costs and increasing yields.²⁵ Yields are expected to increase to 85-95% by 2020. Recovery and reuse of the enzymes also helps to minimize costs, although they do degrade over time.

Acid hydrolysis is the only current cellulose hydrolysis process that does not use enzymes. Acid hydrolysis can only be used in conjunction with an acid hydrolysis pretreatment (2-stage acid process). *Dilute acid hydrolysis*, the oldest cellulosic conversion to ethanol process (the first commercial plant was operating in 1898), is not currently competitive with the enzymatic hydrolysis process. However, *concentrated acid hydrolysis* has significantly higher conversion rates than dilute acid hydrolysis, with sugar yields around 90%. Since acid hydrolysis has high operating costs, minimal acid use and maximal recovery are needed.

Fermentation:

In the fermentation process, microorganisms (bacteria, yeast, or fungi) convert carbohydrates to ethanol in an anaerobic environment. The chemical reactions for conversion of 5-carbon and 6-carbon sugars to ethanol are shown below.



Genetic engineering has been used to generate bacteria and yeast that can ferment both types of sugars (Hamelinck, van Hooijdonk and Faaij 2005). Continued research to provide microorganisms with higher conversion efficiency and resistance would further improve process economics. According to McAloon et al. (2000), the hydrolysis and fermentation steps to produce ethanol from lignocellulosic biomass take about 7 days.²⁶ For comparison,

²⁴ Cellulase enzymes are actually a complex mixture of enzymes, produced by organisms that live on cellulosic material (Hamelinck, van Hooijdonk and Faaij 2005).

²⁵ Key enzyme producers including Genencor International and Novozymes Biotech have both been successful in these developments in recent DOE-supported research.

²⁶ Although long-term projections include reduced time for fermentation. For example, Wooley et al. (1999) estimate a decreasing residence time requirement as technology and microorganisms are further developed, such that fermentation only requires 2 days in the long-term scenario.

the existing mill technology for converting starch to ethanol (not including pretreatment) only takes 2 days.

McAloon et al. (2000) estimate that 39 lb lignocellulosic residue per gallon of fuel ethanol is produced using a dilute acid pretreatment process with corn stover feedstock. The residue typically has a moisture content around 60%. Potential co-products from ethanol production include cell matter, furfural, and acetic acid.²⁷

3.1.2 Performance and cost analysis

Projected cost and performance for ethanol production from the LCE Fermentation/Hydrolysis process has been developed by looking at different pretreatment options for three distinct time periods. The time periods in this analysis include near term (2010), mid term (2015-2020), and long term (2025+). The near term model only includes dilute acid pretreatment and relatively small sizes (25 to 60 MGY). The mid term model uses dilute acid pretreatment for medium size facilities (60 to 100 MGY) and steam explosion for large facilities (>100 MGY). The long term model uses dilute acid pretreatment for medium size facilities (60 to 100 MGY) and liquid hot water (LHW) for large facilities (>100 MGY). The pretreatment technology and size ranges for each time period were selected based on the projections from NREL, Utrecht University, and other institutions (details given below).

This LCE Fermentation/Hydrolysis production model has been derived based on available data from published studies.²⁸ All of the studies used to develop the analysis use enzymatic hydrolysis for cellulose conversion, and a variety of pretreatments for hemicellulose hydrolysis (dilute acid, steam explosion, and LHW). The references use of either poplar or corn stover feedstocks, and all except one use a boiler for power and steam production (there is one that uses biomass integrated gasification combined cycle - BIGCC).

Most of the papers used in the analysis presented a time frame for development of a commercial ready process. The data were grouped by the number of years projected for each technology (i.e. a "2005" technology described by Wooley et al. (1999) was projected to be developed in 6 years). Technologies in the near term include those developed within 5 years, mid term technologies are those developed within 6 to 15 years, and long term technologies are those that are expected to be available after more than 15 years.

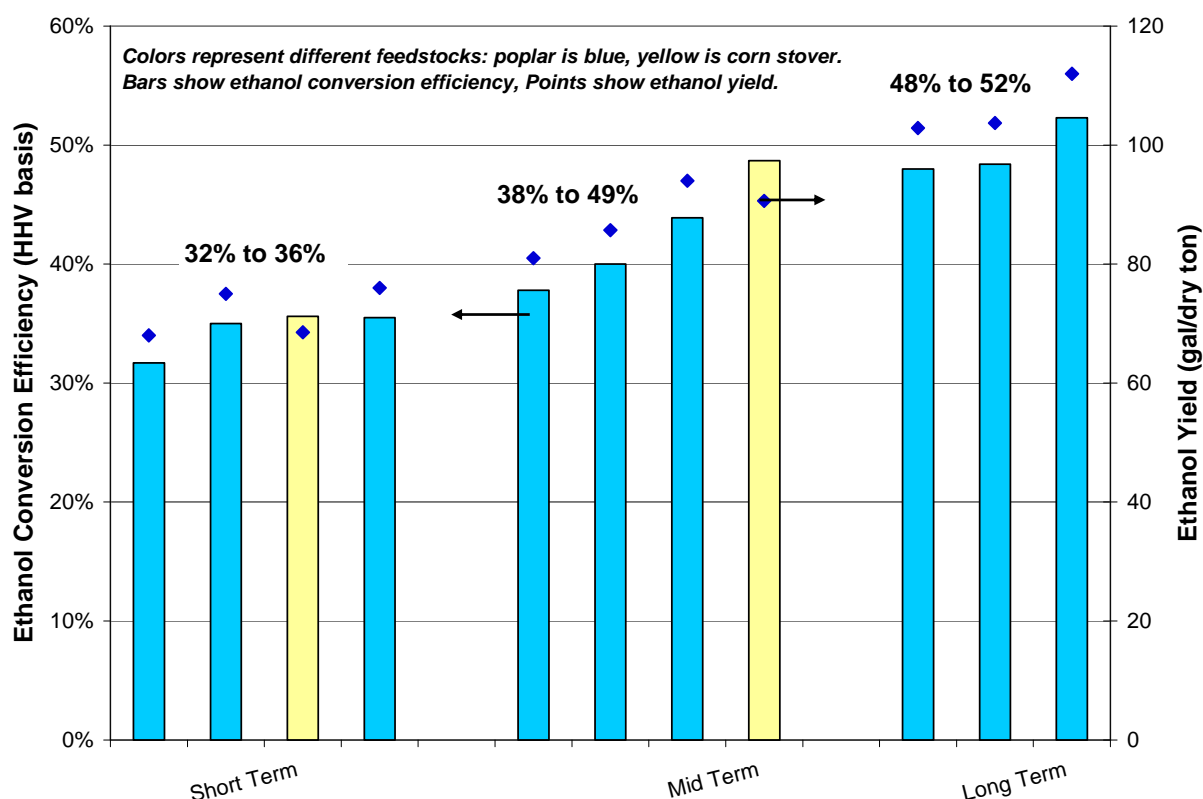
Figure 6 shows the values reported in various studies for ethanol conversion efficiency (bars) and yield (points) for various time periods.²⁹ The conversion efficiency and yield are directly related. The conversion efficiency is calculated as the ratio between the heating value (HHV) of the ethanol produced and the HHV of the feedstock input. The reported values include a variety of pretreatment technologies over a range of projected time periods. There is a clear correlation between ethanol conversion efficiency (or yield) and time period, as the long term processes are projected to approach theoretical conversion efficiencies. This development of production performance will be an important factor in the economic favorability of these technologies.

²⁷ However, as detailed studies have not been made on the development of these co-products they will not be included in this analysis. Electricity is the only by-product from the process evaluated here.

²⁸ Sources: Hamelinck, van Hooijdonk and Faaij 2005; Wooley et al. 1999; Lynd et al. 2005; Aden et al. 2002; and McAloon et al. 2000.

²⁹ Ibid

Figure 6 – Ethanol production efficiency and yield from published data for various time periods



Note: Arrows are used on the figure to reference which axis pertains to each series. The bars represent ethanol conversion efficiency (left axis), while the points show ethanol yield (right axis).

A near term conversion process based on dilute acid hydrolysis with a 35% conversion to ethanol has an overall process efficiency of about 60% including electricity production from lignin and other residuals in a boiler (Hamelinck, van Hooijdonk and Faaij 2005). Improvements in the pretreatment methods and other processes could lead to a 48% ethanol conversion efficiency (or greater), and overall process efficiencies of 68% (Hamelinck, van Hooijdonk and Faaij 2005).

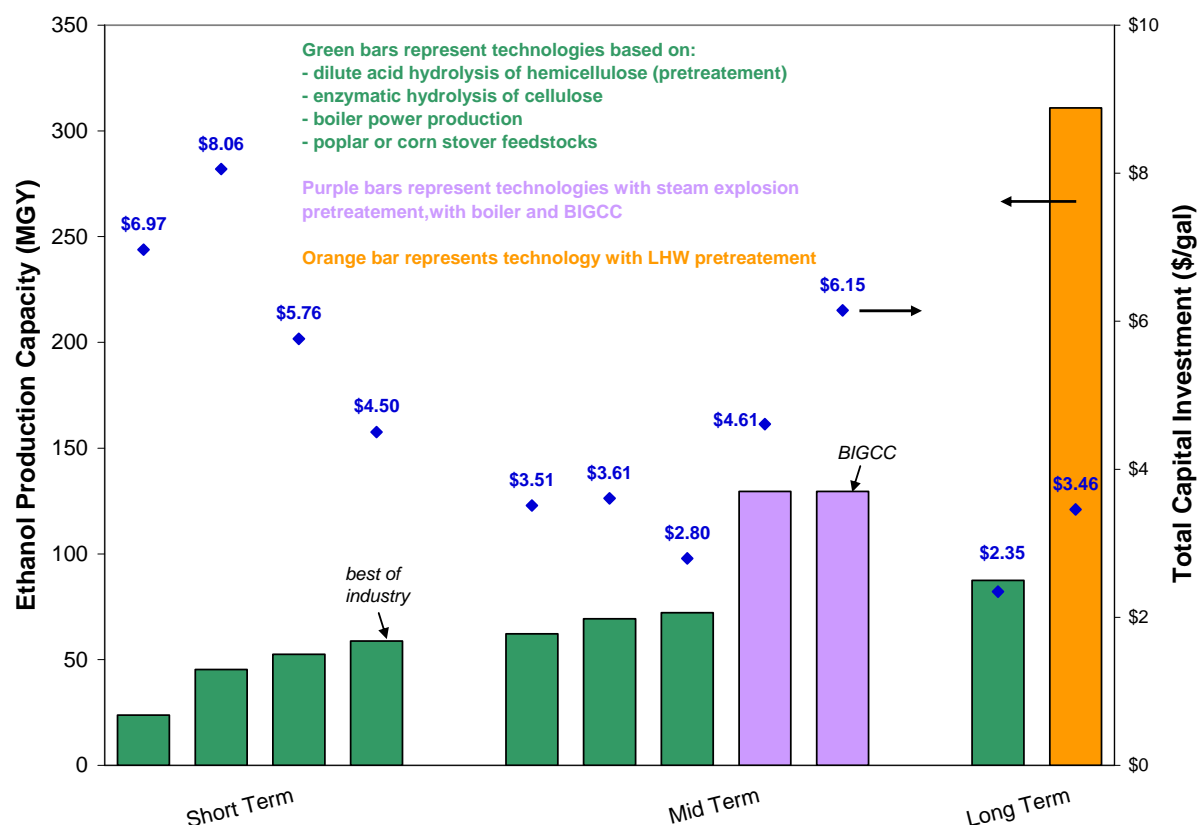
Figure 7 shows the total capital investment in relation to the ethanol production capacity and time period.³⁰ As with the ethanol production performance data discussed above, these values are from a number of studies and include several pretreatment technologies. As there are no commercial lignocellulosic ethanol production facilities in operation, these values have been derived using factored estimation analysis or detailed process design.³¹ The Total Capital Investment (TCI) is generally +/- 30% (or better), and includes the equipment costs, installation, contingency and other indirect costs.³²

³⁰ Ibid (except Lynd et al. 2005).

³¹ Iogen currently has an operational demonstration facility in Canada. A number of companies have demonstration and commercial facilities planned for start-up in the near-term (Solomon, Barnes and Halvorsen 2007).

³² All costs are reported in 2006 US\$, converted using the Chemical Engineering Plant Cost Index (CEPCI) where necessary.

Figure 7 – Ethanol production capacity and total capital investment for various time periods



As expected there is a clear trend towards increased plant sizes with time. The average short term plant size is 45 MGY, while the long term facilities are projected to be about 100 to 300 MGY. Furthermore, the capital cost decreases with time, which corresponds with the economies of scale as plant sizes increase, as well as the availability of lower cost equipment.

In the short term, the only clear outlier with respect to capital cost is the “best of industry” (labeled in graph), which has a TCI of \$4.50/gallon. The outlier for the mid term is the facility which includes gasification for power production instead of a boiler (labeled “BIGCC”). The TCI for this facility is \$6.15/gal, significantly higher than the equivalent facility which uses a boiler (\$4.61/gal). As both of these examples have significantly different technologies from the other modeled facilities, their cost information is not included in the average used for the ethanol production cost model.

The relative costs of the major equipment areas from several studies for near term LCE Fermentation/Hydrolysis facilities using dilute acid pretreatment technology are shown in Table 8. Note that the first two facilities include cellulase production, while the 69 MGY plant does not.³³ Note that at least for the near term, it is more likely that enzymes would be purchased rather than produced.

³³ While some of the LCE Fermentation/Hydrolysis facilities from the published studies used in the analysis include on-site production of cellulase for enzymatic hydrolysis, others assume the cellulase is purchased for use, depending on which option was more economically favorable. The cost of purchased cellulase is included in the consumables portion of the Variable O&M costs.

The power and steam production area is the most expensive, followed by pretreatment. This suggests that lowering the costs of these process steps could be most beneficial in reducing overall capital costs.

Table 8 - Capital cost break-out for LCE Fermentation/Hydrolysis with dilute acid pretreatment

| Plant Size (MGY) | 25 | 52.5 | 69 |
|---------------------------------|----------------------------|---------------------------|-------------------------|
| Major Equipment Area | % of total cost | | |
| Feedstock Handling | 4.0% | 3.6% | 6.6% |
| Pretreatment | 21.9% | 19.5% | 23.6% |
| Saccharification / Fermentation | 10.6% | 9.9% | 8.3% |
| Cellulase Production | 13.3% | 11.5% | 0.0% |
| Distillation / Separation | 10.5% | 9.6% | 19.2% |
| Power / Steam Production | 27.6% | 33.0% | 33.7% |
| Other* | 12.2% | 12.9% | 8.8% |
| <i>Total</i> | 100.0% | 100.0% | 100.0% |
| Source | <i>McAloon et al. 2000</i> | <i>Wooley et al. 1999</i> | <i>Aden et al. 2002</i> |

* Other includes wastewater treatment, storage, and utilities

Outline for model analysis

Due to the complexity of the LCE Fermentation/Hydrolysis process, it was necessary to use a number of assumptions to generate a simple and useable model. The cost and performance model was developed as follows:

- Hemicellulose hydrolysis (pretreatment) varies based on facility size and time period
 - The only available technology for short term (2010) is dilute acid pretreatment, in the size range from 25 to 60 MGY.
 - Mid term (2015-2020) pretreatment technologies are dilute acid for medium size facilities (60 to 100 MGY), and steam explosion for large facilities (> 100 MGY).
 - Long term (2025+) pretreatment technologies include dilute acid for medium size facilities, and Liquid Hot Water (LHW) for large facilities.
 - It is assumed that small facilities (< 60 MGY) are not constructed after the short term time period.
- Cellulose hydrolysis step is performed using enzymatic hydrolysis (with cellulase) for all modeled processes.
- Cellulase is produced on-site for most scenarios used in this analysis, although in some cases it is purchased. The selection for on-site production versus purchase of cellulase follows the choices reported in the literature for each technology option. On-site cellulase production will affect O&M costs, capital costs, and overall ethanol yields, as a fraction of the sugar stream is diverted to feed the microorganisms. Purchased cellulase only affects O&M costs, and is generally more expensive overall.
- Electricity and steam are produced with a boiler and turbogenerator from residual solids (lignin, unconverted holocellulose, cell mass), and unconverted sugars which are partially dried to a syrup.
- Electricity is the only co-product produced from the process, and is valued at 5.7¢/kWh based on the average industrial electricity cost from Energy Information Administration (EIA 2006). Note that EIA's projections for industrial electricity for the 2010 to 2030 time period is 5.8¢/kWh, only slightly higher than current value (EIA AEO 2007).

Selection of feedstock type is a critical component of the model, as the composition of the biomass affects the ethanol production. In particular, the hemicellulose, cellulose, and lignin content of the feedstock affect yield (gallons of ethanol produced per dry ton of feedstock). The pretreatment technology also has a significant effect on the ethanol yield,

as each technology has a different conversion factor. The feedstock composition and yield in turn affect the material available for power production – a portion of the unfermented sugars are dried and the resulting syrup is fired in the boiler along with the lignin and cell mass. In addition, different feedstocks will have different costs and availability.

The conversion rates of hemicellulose and cellulose into sugars and the fermentation of sugars into ethanol are technology dependent. These conversions also increase with time, such that for each pretreatment process the conversion rates in the long term are significantly higher than in the short term rates. Conversion rates for saccharification of cellulose and hemicellulose and fermentation of the glucose and other sugars are expressed as a percent of theoretical conversion for each technology available in each time period, as shown in Table 9.

Table 9 - Conversion rates for LCE Fermentation/Hydrolysis

| Conversion (% of theoretical) (1) | | | | | |
|-----------------------------------|---|-----------------------|--------------------------|-----------------------|--------------------------|
| | Short Term | Mid Term | | Long Term | |
| Pretreatment Technology | Dilute Acid | Dilute Acid | Steam Explosion | Dilute Acid | LHW |
| <i>Saacharification</i> | | | | | |
| Hemicellulose | 83% | 85% | 55% | 85% | 93% |
| Cellulose | 75% | 85% | 93% | 90% | 98% |
| <i>Fermentation</i> | | | | | |
| Xylose & other sugars (2) | 86% | 90% | 85% | 95% | 94% |
| Glucose (2) | 93% | 94% | 93% | 95% | 94% |
| Source | Hamelinck et al. 2005, Wooley et al. 1999 | Wooley et al. 1999 | Hamelinck et al. 2005 | Wooley et al. 1999 | Hamelinck et al. 2005 |

1) Average conversion estimates based on poplar feedstock and the reported pretreatment technologies. The advanced technologies for mid and long term pretreatments may not be suitable for all feedstocks.

2) Cellulose is converted to glucose during saacharification, while hemicellulose is converted to xylose and other sugars

The theoretical ethanol yield from biomass feedstocks is derived based on the dry weight of fraction of C-5 and C-6 sugars. Theoretical yield is equal to the total weight percent of C6 sugars times 172.82, plus the total weight percent of C5 sugars times 176.87 (U.S. DOE EERE Biomass Program 2006).³⁴ The calculation for actual ethanol yield is very similar, except that the percent of theoretical conversion for each step is taken into account.

The electricity production depends on the amount of material available for boiler fuel and the steam use, which are related to the facility type. The dilute acid distillation process and LHW pretreatment have high steam requirements which implies lower electricity production. Facilities with higher conversion rates will have fewer unconverted sugars to use a boiler fuel.

Operations and Maintenance (O&M) costs are separated into fixed and variable components. The fixed O&M includes annual costs which are a function of the plant's capacity not its annual ethanol production. These costs include labor, maintenance, taxes and insurance. Variable O&M costs include raw materials (such as sulfuric acid and lime for dilute acid pretreatment, ammonia and corn steep liquor for cellulase production, or purchased

³⁴ These factors are derived from the following conversions: 1.11 pounds of C6 sugar per pound of C6 polymeric sugar, 1.36 pounds of C5 sugar per pound of C5 polymeric sugar, each pound of sugars can produce a maximum of 0.51 pounds of ethanol, and there are 6.55 pounds of ethanol per gallon.

cellulase if it is not made in the process) and solids disposal (i.e. ash and gypsum). The O&M values used in the model are derived from published studies for each technology type.

The key values and assumptions used to calculate the mid-term (2015) technology analysis and economic performance model for LCE Fermentation/Hydrolysis are described below. An example of the detailed analysis for all periods is shown in Appendix G.

- Feedstock type and input quantity are variables. Applicable feedstocks include corn stover, wheat straw, switchgrass, poplar (hardwood) and pine (softwood), and portions of the MSW stream (see Appendix f for details on biomass from MSW). These feedstocks were selected based on available composition data, and can be used to represent a range of wood and agricultural residues.
- The applicable size range for facilities in this time period is 60 to 150 MGY, based on facility sizes from sources (Hamelinck, van Hooijdonk and Faaij 2005; Wooley et al. 1999; Aden et al. 2002). Facilities with capacities from 60 to 100 MGY are modeled to use dilute acid pretreatment, while larger facilities (> 100 MGY) use steam explosion pretreatment.
- Ethanol yield is based on the feedstock composition and pretreatment technology. Biochemical compositions (hemicellulose, cellulose, and lignin percentages) for the feedstocks are based on data from U.S. DOE EERE Biomass Program (2006) and Hamelinck, van Hooijdonk and Faaij (2005), as shown in Table 7 above. The actual yields are calculated based on projected conversion of hemicellulose and cellulose to sugars during saccharification, and the fermentation of these sugars to ethanol. The projected conversion rates for each technology are given in Table 9 above. Yield data for various feedstocks are shown in Table 10.
- The total conversion efficiency (HHV) is 38-41% for dilute acid and 36-40% for steam explosion, depending on the feedstock type. Corn stover and pine have the highest conversion rates of the selected feedstocks.
- Capital cost data is based on average of results from (Hamelinck, van Hooijdonk and Faaij 2005; Wooley et al. 1999; Aden et al. 2002; and McAloon et al. 2000).³⁵ Each pretreatment technology has a different base capital cost, which is scaled by facility size using a typical scaling factor for biofuels of 0.8. The base cost for a mid term dilute acid facility is \$3.31/gal for a 68 MGY facility, and the cost for a 130 MGY steam explosion facility is \$4.61/gal.
- Net electricity production is based on heat and power production with partially dried solid residuals (lignin, cell mass) and unfermented sugars (dried to a syrup), fired in a boiler. The amount of electricity generated is dependent on the pretreatment technology. For the mid-term period, electricity production for the dilute acid process is about 2 kWh per gallon of ethanol produced, following Wooley et al. 1999, while steam explosion pretreatment gives about 6.3 kWh/gal based on data from Hamelinck, van Hooijdonk and Faaij (2005). Note that the net electricity production for the steam explosion pretreatment is much higher than other technologies as this method has lower steam requirements. Furthermore, cellulase is not made on site in the steam explosion case, which means a larger quantity of non-fermented sugars and residuals are available for power generation.
- Annual water consumption is calculated based on the annual feedstock input, using data from Wooley et al. (1999) for near-term dilute acid process. Potential changes in water consumption for different time periods or for different pretreatment processes are not included in this analysis.

³⁵ Note that the BIGCC case and best of industry are excluded from capital cost estimates, as they are not representative of an average facility.

- The CO₂ production stream is estimated by assuming that the amount of CO₂ produced is approximately 96% of the weight of ethanol produced, since the theoretical max yield is 51% ethanol and 49% CO₂ (by weight) from sugars.
- Annual Fixed and Variable O&M costs are calculated based on data in published studies. Mid-term dilute acid O&M costs are based on Wooley et al. (2005), and steam explosion data is from Hamelinck, van Hooijdonk and Faaij (2005). Variable O&M includes raw materials (sulfuric acid & lime for dilute acid, ammonia & CSL for cellulase production, other chemicals) and solids disposal (i.e. ash and gypsum). Fixed O&M includes labor, maintenance supplies, insurance and property taxes.
- The co-product credit for electricity is based on the net electricity available for export and a current electricity value of \$0.057/kWh, based on the 2005 average wholesale price of electricity for industrial sector from EIA (2006).
- The levelized non-feedstock production cost for ethanol is calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Table 10 - Feedstock Composition and Yield Data

| Feedstock | Yield | Actual Yield (gal/ton) (2) | | | | |
|-------------------------------|---|----------------------------|------------------------|----------------------------|-------------------------|-----------------|
| | Theoretical Max Yield (1) (gal/dry ton) | Short Term (Dilute Acid) | Mid Term (Dilute Acid) | Mid Term (Steam Explosion) | Long Term (Dilute Acid) | Long Term (LHW) |
| Agricultural Resources | | | | | | |
| Corn Stover | 107.6 | 72.6 | 80.6 | 77.3 | 86.0 | 82.0 |
| Wheat Straw | 98.8 | 69.3 | 76.8 | 73.6 | 82.0 | 77.9 |
| Grass Resources | | | | | | |
| Switchgrass | 99.7 | 70.0 | 77.4 | 74.4 | 82.6 | 78.7 |
| Woody Resources | | | | | | |
| Hybrid Poplar | 109.9 | 76.9 | 85.9 | 81.2 | 91.9 | 98.6 |
| Pine | 115.1 | 80.9 | 90.2 | 83.7 | 96.4 | 103.5 |
| Other Resources | | | | | | |
| Yard Trimmings / Green Waste | 108.0 | 63.9 | 70.0 | 68.6 | 74.4 | 73.4 |
| Mixed Paper | 116.2 | 76.8 | 86.0 | 83.5 | 92.1 | 98.8 |

1) Maximum yield values based on sugar content, calculated from EERE's Theoretical Ethanol Yield Calculator (http://www1.eere.energy.gov/biomass/ethanol_yield_calculator.html)

2) Actual yields calculated based on average conversion values for all feedstocks in short term, and wood for all terms. Agricultural and grass feedstock yields for advanced mid and long term pretreatment technologies based on average increase of wood feedstock yields from short to mid term (4.4% of theoretical), as the conversion technology information does not apply to these materials, but other comparable advanced technologies may apply.

Model Results:

Some of the key cost and performance results for projected mid-term LCE Fermentation/Hydrolysis processes are shown in Table 11 and Table 12. These tables show results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs for each pretreatment technology. The analysis shows that in general the non-feedstock production cost of ethanol for dilute acid is lower than the steam explosion. Although the steam explosion process generates significantly more electricity, it also has much higher capital and O&M costs than dilute acid. However, steam explosion has environmental and other benefits that are not accounted for in this simple economic analysis that may encourage development of these facilities.

Table 11 - Example Cost and Performance - Mid Term Dilute Acid LCE Fermentation/Hydrolysis

| Sample Model Results - LCE Fermentation/Hydrolysis w/ Poplar (dilute acid) | | |
|---|---------------|---------------|
| Feedstock Input (dry ton/yr) | 700,000 | 1,160,000 |
| Ethanol Yield (MGY) | 60 | 100 |
| Electricity Production (MWh/yr) | 120,800 | 200,200 |
| CO ₂ Stream (ton/yr) | 190,400 | 315,500 |
| Water Consumption (1000 gal/yr) | 373,900 | 619,600 |
| Capital Cost (Million \$) | \$203.4 | \$304.7 |
| Annual O&M Costs (Million \$/yr) | \$17.7 | \$29.3 |
| By-Product Credit (Million \$/yr) | \$6.9 | \$11.4 |
| Non-feedstock Production Cost (\$/gal) | \$0.60 | \$0.56 |

Table 12 - Example Cost and Performance - Mid Term Steam Explosion LCE Fermentation/Hydrolysis

| Sample Model Results - LCE Fermentation/Hydrolysis w/ Poplar (steam explosion) | | |
|---|---------------|---------------|
| Feedstock Input (dry ton/yr) | 1,250,000 | 1,850,000 |
| Ethanol Yield (MGY) | 101 | 150 |
| Electricity Production (MWh/yr) | 642,160 | 950,390 |
| CO ₂ Stream (ton/yr) | 321,500 | 475,800 |
| Water Consumption (1000 gal/yr) | 667,700 | 988,200 |
| Capital Cost (Million \$) | \$491.2 | \$672.1 |
| Annual O&M Costs (Million \$/yr) | \$39.6 | \$58.6 |
| By-Product Credit (Million \$/yr) | \$36.6 | \$54.2 |
| Non-feedstock Production Cost (\$/gal) | \$0.62 | \$0.58 |

Hamelinck, van Hooijdonk and Faaij (2005) estimate the mid term ethanol production cost from poplar using steam explosion pretreatment is about \$1.42 per gallon for a facility with 600,000 dry ton per year input. Using similar feedstock cost (\$57 per dry ton) and facility size gives a comparable ethanol production cost from the model, about \$1.27/gal. For further comparison, Aden et al. (2002) estimate an ethanol production cost of \$1.35/gal for a 69 MGY dilute acid facility using corn stover feedstock.³⁶ The feedstock cost from the study is about \$38/dry ton. This gives an ethanol production cost from the LCE model of \$1.04/gal for this configuration, slightly lower than the projection from Aden et al. Different economic analysis methods may contribute to this variation.

3.2 Lignocellulosics to ethanol: Thermochemical Conversion and Fermentation

In the Lignocellulosics to Ethanol via ThermoChemical conversion and Fermentation (LCE Fermentation/TC) process, gasification is used to convert biomass feedstocks into a syngas rich in CO and H₂ which is then fermented to produce ethanol. As this technology is based on gasification, it has a wide range of feedstock flexibility. Furthermore, unlike the LCE Fermentation/Hydrolysis process, the lignin part of the biomass directly contributes to syngas production. Gasification requires dried and sized material, so biomass feedstocks must be pretreated before use.

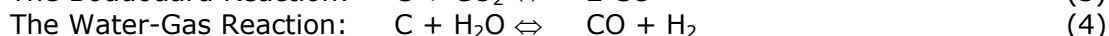
³⁶ Costs have been converted to 2006\$ using CEPCI.

3.2.1 Technology Description

Gasification Overview

Gasification is the process of converting a solid fuel into a gaseous fuel. Gasification is appealing for biomass because a wide variety of materials can be gasified, some of which could be difficult or undesirable to combust directly (such as sludges and hazardous wastes). There is also a lot of flexibility in the end use of the gas. Provided it is cleaned of contaminants, a gaseous fuel can be used to make biobased liquids for liquid fuels and chemicals, or it can be used in high-efficiency power generation systems such as a gas turbine-combined cycle and fuel cells.

The two main reactions occurring during gasification are as follows:



These reactions require heat input. In simple gasifiers the heat is drawn from the combustion of the pyrolysis gases formed during gasification, while in more complex gasifiers it is provided externally. These reactions produce more syngas from the biomass and typically leave little unreacted carbon in the ash.

In general, the carbon conversion in a fluidized bed gasifier increases as the temperature increases. Furthermore, a recent study has shown that the feedstock is an important factor in the conversion behavior (Moilanen 2006). For example, pine wood has high carbon conversion at relatively low temperatures, whereas pine bark has relatively low conversion efficiency even at higher temperatures. Additionally, straw has high conversion efficiencies above 850°C, but ash sintering causes operational problems.³⁷ This study also demonstrated that the reactivity of the biomass (gasification rate) is dependent on the biomass fuel properties, the gasifying agent, and the temperature and pressure conditions during reaction (Moilanen 2006). Some of these effects seem to be related to the behavior of the ash-forming substances during gasification, which are difficult to characterize.

The main challenges facing biomass gasification are as follows:

- The high moisture content of un-dried woody biomass;
- Commercial availability of gas clean-up technologies that are sufficiently robust and effective to allow the product gas to be used in a turbine or engine; and
- The relatively high cost of gasification equipment, which is currently built on a custom basis.

Furthermore, gasification for power generation is currently a challenge as there is limited industry experience in using biomass gasifiers in advanced power generation cycles.

There are currently two principal types of gasifiers: fixed bed and fluidized bed. The former is typically simpler, less expensive and produces a low BTU syngas. The latter is more complicated, more expensive and produces a syngas with a higher heating value.

In fluidized bed systems, steam and/or air are passed through a particle bed and cause the bed to levitate (i.e., become “fluidized”). Fuel is added to the bed which is often composed of inert heat transfer media such as sand or dolomite. Other materials such as alumina may be used to further enhance the heat transfer. Fluidized bed devices are noted for their high energy output (per area of bed) and fuel flexibility.

³⁷ Note that these experimental results can be used to describe the characteristics of the fuel, but conversion in a gasifier will be affected by many additional factors, such as the partial pressure of the gasification agent and the size distribution of feedstock particles.

At present, there are numerous small, fixed-bed gasifiers in use around the world. These units typically convert wood and agricultural residues into a low-BTU gas for use in process heating or district heating. These applications have developed as alternatives to traditional boilers. There has been a significant effort to develop large-scale (>10 MW_e) gasifiers for use in a biomass integrated gasification combined cycle (BIGCC). These efforts have not yet resulted in a sustainable commercial operation.

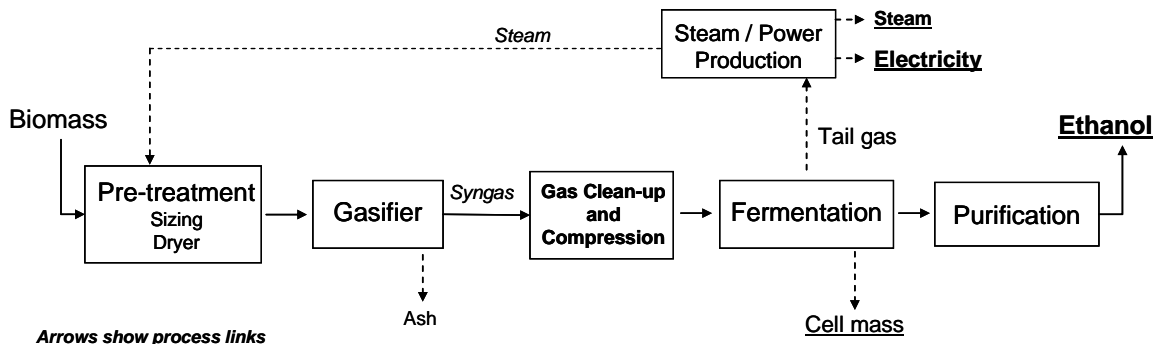
Syngas Fermentation to Ethanol

Figure 8 shows the general arrangement for the LCE Fermentation/TC process. After gasification, the syngas is conditioned and compressed for fermentation. In the fermentor, anaerobic micro-organisms ferment the syngas to produce ethanol. Although these organisms prefer CO to H₂, the syngas can contain any mixture of these gases for conversion. The chemical reactions for ethanol production for this process are as follows:³⁸



Typical CO conversions reported in literature for laboratory scale fermentation are ~90%, while conversions for H₂ only about 70% (Spath and Dayton 2003).

Figure 8 – Schematic diagram showing LCE Fermentation/TC process



Source: Diagram based on figure in (Spath and Dayton 2003)

Unlike the by-products of corn ethanol production, the cell mass produced during fermentation of syngas is not currently approved for animal feed (Spath and Dayton 2003). An alternate use for the material is to recycle it to gasifier for re-processing. The exhaust gas produced from fermentation can be used to produce steam and power. The steam can be used to dry the biomass feedstock during pretreatment.

Theoretically, all the chemical energy in biomass can be converted to syngas. However, not all of the syngas will be converted to ethanol. Estimates of the ethanol yield produced from biomass with this process can be estimated using the following equation, assuming all of the syngas produced by the gasifier (after conditioning) goes to fermentation:³⁹

$$\text{Ethanol(MGY)} = \frac{\text{Feed}_{in}(\text{ton/day}) \times \text{HHV}(\text{Btu/lb}) \times \eta_{\text{gasification}} \times C}{1.5 \times 10^5}; \quad (7)$$

³⁸ Source: (Spath and Dayton 2003)

³⁹ Source: (Spath and Dayton 2003)

Where Feed_{in} is the rate of feed input to the gasifier in dry tons per day;
HHV is the higher heating value of the feedstock;
 $\eta_{\text{gasification}}$ is the cold gas efficiency of the gasifier and gas conditioning; and
C is the conversion rate of CO and H₂ to ethanol, as a fraction of theoretical

Projected yields for LCE Fermentation/TC process are 70-105 gallons of ethanol per dry ton of feedstock (Spath and Dayton 2003). This is lower than the yields from current grain ethanol production technologies, which are around 100 gal/ton. Spath and Dayton (2003) note that this is a result of losses from gasification, gas conditioning, and fermentation steps which counteract the benefits of direct utilization of the lignin component of the biomass.

University of Arkansas and Bioengineering Resources Inc (BRI) have been working on this technology since the 1980's. BRI has a patented fermentation bacteria for the process (Bioengineering Resources 2007), and is part of a team that was recently selected to build a demonstration facility with funding from DOE (U.S. DOE Feb. 2007). Several other universities have also been involved in research efforts. However, as yet this technology does not seem to be near commercialization, and there is not sufficient information about the cost and performance for a detailed model. As such, the syngas fermentation process for ethanol production will not be further characterized here.

3.3 Lignocellulosics to Middle Distillates: Fischer Tropsch

Fischer Tropsch (FT) synthesis is a method of generating hydrocarbons from gaseous mixtures of CO and H₂ using a metal catalyst. This process has been known since the early 1900's, and the first commercial plant was built by Sasol in South Africa in 1955 (using coal syngas). FT synthesis produces a range of hydrocarbon products, and a large amount of heat from the highly exothermic reaction. The distribution of products generated from FT synthesis is affected by reaction temperature and pressure, input syngas composition, and catalyst type and composition. Among many other products, FT can be used to generate diesel and gasoline fuels. Although the FT process is relatively well characterized, there has not been a lot of analysis or testing with biomass feedstocks.

3.3.1 Technology Description

The general steps for the LignoCellulosics to Middle Distillates – Fischer Tropsch (LCMD – FT) process are shown in Figure 9. As typical for gasification processes, the biomass must first be sized and dried for use (see detailed gasification description in Section 3.2.1 above). After gasification, impurities such as particulates, sulfur, and tar must be removed from the raw syngas. The amount of gas clean-up required will depend on the type of process and economics. However, this is a very important step in the LCMD - FT process, as the FT catalysts (Iron (Fe) or Cobalt (Co)) are extremely sensitive to chemical poisoning by sulfur, as well as by halides and nitrogen.

After clean-up, the syngas is then sent to *either* a Low Temperature (200-240°C) *or* a High Temperature (300-350°C) FT reactor (Spath and Dayton 2003). Reactor pressures are generally 145-580 psi (Spath and Dayton 2003). Low temperature synthesis typically uses Co catalysts, and is used to maximize the diesel fraction of the product, while high temperature synthesis uses Fe catalysts and maximizes gasoline production.

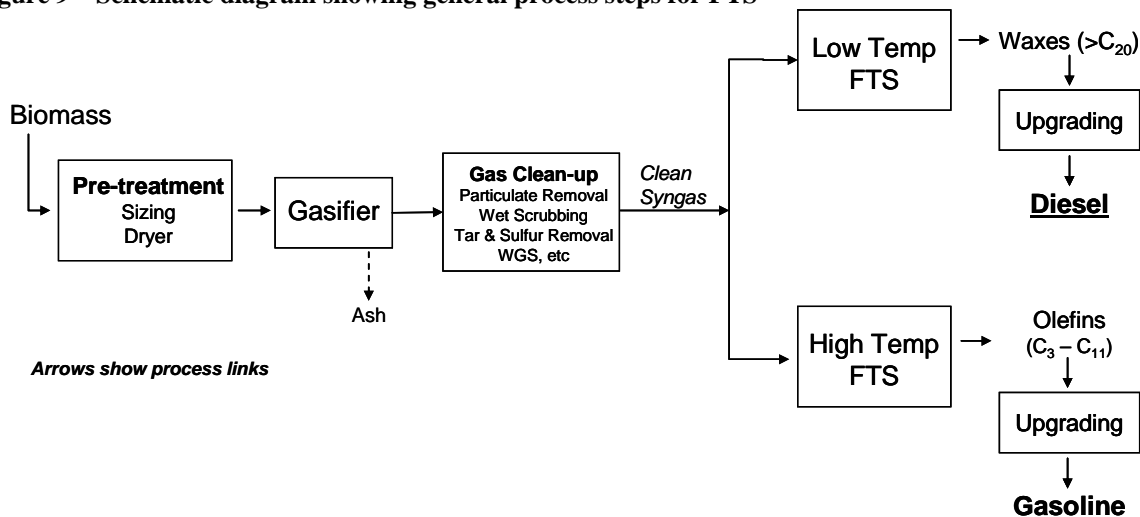
There are four types of reactors that have been used commercially, including:

- **Multi-tubular fixed bed**, also known as the ARGE reactor. Has conversion efficiencies around 70%, and the products are about 50% wax)
- **Circulating fluidized bed**, such as the original Synthol reactors which produce high fractions of gasoline and light olefins. These reactors are complex, and the circulating bed material causes considerable erosion which leads to high O&M costs. They have reached design limit for capacity at 7,500 bpd.
- **Fixed fluidized bed**, or Sasol Advanced Synthol reactor, which has replaced the older Synthol reactor. It is half the size and cost for same capacity as the Synthol reactors, and has lower operating costs. Size limit is 20,000 bpd.
- **Fixed slurry bed** reactors, which are low temperature reactors optimized for wax production. They cost less than tubular reactors, but have only recently been used for commercial FTS. Separating the wax from the catalyst can be a challenge.

Temperature control is critical in the reactor, as it affects the product distribution and can harm the catalyst. In addition to the chemical poisoning mentioned above, catalysts lose activity due to sintering, carbon deposition, and oxidation. Furthermore, reactor temperatures above 400°C encourage methane formation, an unwanted gaseous product (Spath and Dayton 2003).

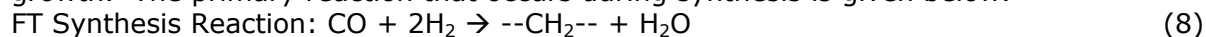
The off-gas produced from the FT reactor can be diverted and recycled back through the reactor to generate additional hydrocarbon products, or it can be used to generate power/steam (not shown in the figure). Product liquids generated from FT synthesis are upgraded using a combination of hydrotreating, hydrocracking, hydroisomerization, and separated.

Figure 9 – Schematic diagram showing general process steps for FTS



Based on figure in (Spath and Dayton 2003)

Fischer Tropsch synthesis is characterized as a polymerization reaction with stepwise chain growth. The primary reaction that occurs during synthesis is given below.



The --CH₂-- is a building block for hydrocarbon chains. A secondary reaction occurs when Fe catalysts are used, Water-Gas Shift (WGS).



The combination of these reactions gives the net overall reaction for Fe catalyzed FT synthesis:



The syngas entering the reactor needs to have a H_2/CO ratio of 2.15 for a cobalt catalyst, and 1.7 for an iron catalyst (lower because of the WGS reaction). The composition of the syngas can be altered by changing the gasification conditions, or by processing the syngas after clean-up via WGS, CO_2 removal, or methane reforming (which converts methane with steam into CO and H_2).

It is clear that FT catalysts are a very important part of the synthesis. Co catalysts tend to last longer than Fe, and they have better carbon conversion to products, since there is no major impact from the secondary WGS reaction. Additionally, Co doesn't require as high of pressures as Fe, which equates to lower operating costs. However, Fe is attractive because it is much less expensive than Co (approximately 200 times cheaper).

The product selectivity from FT synthesis is based on the catalyst's ability to encourage chain growth versus chain termination. Regardless of operating conditions, the Fischer Tropsch reaction always produces a range of olefins, paraffins, and oxygenated compounds like alcohols, aldehydes, acids, and ketones. Only methane and paraffin wax can be produced with very high selectivity (Spath and Dayton 2003). The maximum gasoline product fraction selectivity is 48%, while for diesel the maximum is about 40% (Spath and Dayton 2003).

Technology Status

There has been a lot of research and development of FT synthesis technologies. Sasol has several commercial FT plants in South Africa that use coal and crude oil feedstock. Sasol's first FT plant was built in 1955 and because of its success two more were built in 1980's. These facilities are currently using Sasol Advanced Synthol reactors, which have much lower O&M costs than the original Synthol reactors.

Two other plants became operational in the 1990's, both of which use natural gas feedstock. The Moss gas plant is a Fischer Tropsch facility in South Africa that produces 1.1 million tons/yr of FT products (Spath and Dayton 2003). Shell commissioned a plant in Malaysia in 1993 which produces 0.6 million tons per year of FT products.

3.3.2 Performance and cost analysis

The LCMD - FT performance and cost model has been developed primarily based on recent work Antares performed for US DOE and NREL (Antares 2003a, Antares 2003b, and Antares 2003c). In this analysis, Antares developed process models, capital costs, and O&M costs for several FT systems with varied size and technology types. One of the commercial scale systems modeled previously was selected for use as the basis for mid-term (2015) LCMD - FT technology in this analysis.

The defining characteristics of the configuration are the gasifier, FT reactor and steam reforming system. The selected configuration uses an air-blown, indirect, atmospheric gasifier, which is represented by the BCL/FERCO gasifier (abbreviated as BCL). After cleaning and conditioning, the compressed syngas is processed in a steam methane

reformer (SMR) to convert much of the hydrocarbons to carbon monoxide and hydrogen.⁴⁰ The syngas then undergoes a shift reaction to produce an H₂ to CO ratio of about 2:1. This configuration uses two (low temperature) fixed bed FT reactors with Co catalysts. Each reactor has a CO conversion of about 42%, so that the overall conversion is 66%. Two reactors are used in this process to increase the yield of FT-liquids, as recycling is not an effective option for air-blown gasifiers. The unconverted syngas is used for electricity and heat production in a combined cycle power plant. The FT-liquids produced in the reactor are hydrocracked and undergo primary distillation before they are sent to an off-site refinery. The final products are distillate and naptha. A diagram showing the major components and process flows for this system is shown in Figure 10.

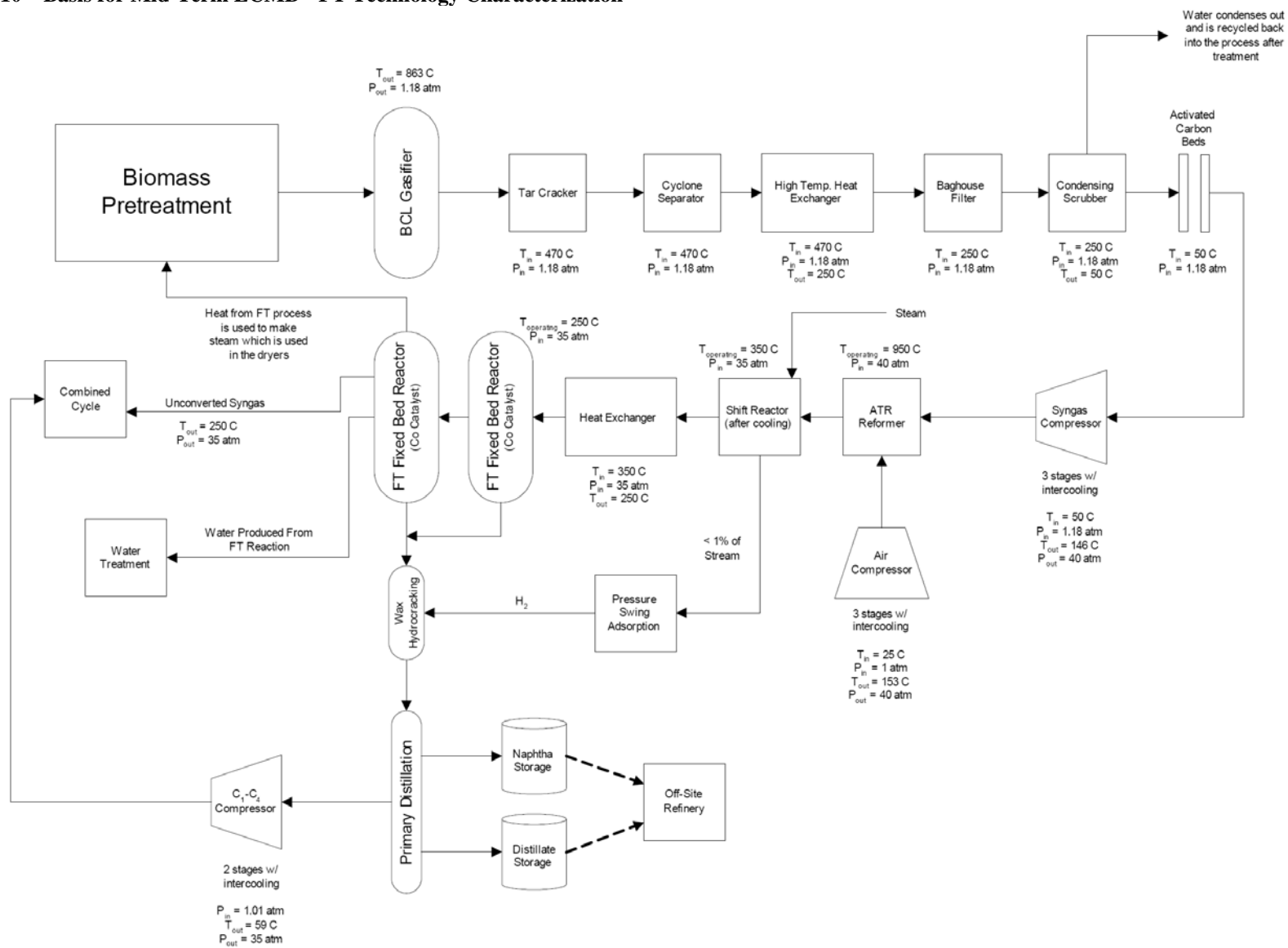
In general, the total capital investment for cases involving SMR is greater than that for the cases involving ATR. However, it is important to realize that the reformer choice affects O&M costs. Relative to ATR, the use of SMR increases FT liquids production and boosts exportable power revenues, thus resulting in lower total O&M costs on a \$/bbl basis.

The overall process efficiency of this system including electricity and steam production is 58% on LHV basis (Antares 2003a). For this configuration of the LCMD - FT process, about 73% of the energy produced is embodied in the FT liquids (C5+), and the remainder is exportable power.⁴¹

⁴⁰ Note that although SMR has higher capital costs than autothermal reforming (ATR), it is also more efficient. ATR is currently the only practical reformer to use for biomass-derived syngas, but it is expected that STR reforming could be available for use by the mid-term period. An optimistic view of SMR is used in the model.

⁴¹ BIG-FT stands for Biomass Integrated Gasification – Fischer Tropsch.

Figure 10 – Basis for Mid-Term LCMD - FT Technology Characterization



Outline for model analysis

The key values and assumptions used to calculate the advanced technology analysis and economic performance model for Fischer Tropsch synthesis are described below. An example of the detailed analysis is shown in Appendix G.

- Feedstock type and input quantity are variables. Applicable feedstocks include woody and agricultural residues.
- This model is applicable for facility sizes with greater than 185,000 dry ton biomass input per year, based on Tijmensen, Faaij and van Hardeveld (2002).
- Yield of FT liquids are calculated based on the LHV conversion efficiency. As different feedstocks have distinct heating values, the feedstock choice affects the amount of FT liquids produced. The conversion efficiency for distillates and naphtha have been calculated in Antares 2003a. For the selected system, LHV efficiencies for the products are 30.4% for distillates and 12% for naphtha.⁴² This leads to a total energy conversion efficiency of 42% for the process.
- The FT liquids produced from this configuration have estimated heating values (LHV) of 125,684 Btu/gal for the distillates and 115,263 Btu/gal for the naphtha.
- The amount of electricity generated is calculated to be about 560 kWh per barrel of FT liquids produced, following Antares 2003a.
- Annual water consumption is equivalent to about 33,500 thousand gallons for a facility with a feedstock input of 700,000 dry ton per year, based on data from Antares 2003a.
- Capital cost data is based on values for a commercial facility with a biomass input of 80 dry ton per hour (about 700,000 ton/yr) from Antares 2003c. The scaling factors are 0.74 for a biomass input from 185,000 – 750,000 dry ton/yr, and 0.91 for a biomass input greater than 750,000 dry ton/yr, following (Tijmensen, Faaij and van Hardeveld 2002).⁴³
- Annual non-labor fixed O&M costs are calculated as a percentage (2.3%) of the capital cost, following Antares 2003a. Maintenance and repair accounts for 2% of the cost based on capital investment, and the operating supplies are estimated to be 15% of the total maintenance cost.
- Labor costs are calculated based on the number of people required to operate the facility, with an assumed current labor rate of \$82,000 per year per employee (including fringe and benefits). The base labor requirement for a facility with 700,000 ton/yr biomass input is \$5,660,000/yr. This cost is scaled by facility size using a scaling factor of 0.25, as in (Tijmensen, Faaij and van Hardeveld 2002).
- Annual Variable O&M costs include steam, water, wastewater treatment, ash and catalyst disposal, consumables, and FT catalysts. These are calculated on a \$/gallon basis of the FT liquid products, following Antares 2003a.
- The co-product credit for electricity is based on the net electricity available for export and a current electricity value of \$0.057/kWh (based on the 2005 average wholesale price of electricity for industrial sector from EIA 2006). The co-product credit for naphtha is estimated to be \$63/bbl, based on average market data over the past year from Plastemarte (2007).⁴⁴
- The levelized non-feedstock production cost for the FT distillates are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

⁴² These yields were calculated using the Anderson-Schulz-Flory (ASF) theory to model the FT product distribution (Antares 2003a). For the cobalt catalyst the chain growth probability is assumed to be 0.93, which corresponds to a C5+ selectivity of 94.9%.

⁴³ "When scale is between 100 and 400 MWth the overall scaling factor for the entire plant (with respect to overall total investment costs) is approximately 0.74. When capacities go beyond 400 MWth, the average scaling factor increases to 0.91" (Tijmensen, Faaij and van Hardeveld 2002, p. 143)

⁴⁴ The market value of naphtha ranged from \$580 to \$755 per ton over the past year (Plastemarte 2007).

Model Results

Some of the key cost and performance results projected for LCMD - FT are shown in Table 13. This table shows results for the range of applicable facility sizes based on selected input quantities, and illustrates the high and low end costs. The low end feedstock input shows the minimum applicable facility size, and the high end was selected to show a reasonable range of sizes and costs. In general, the analysis shows that the non-feedstock production cost of distillates from FT synthesis is fairly expensive. However, the by-product credit is a critical factor and can have a significant effect on the production costs.

Table 13 - Example Cost and Performance for Fischer Tropsch Facility

| Sample Model Results - Fischer Tropsch Synthesis w/ Poplar | | |
|---|---------------|---------------|
| Feedstock Input (dry ton/yr) | 185,000 | 1,500,000 |
| FT Distillate Yield (MGY) | 7.5 | 61 |
| Naptha Yield (MGY) | 3.2 | 26 |
| Electricity Production (MWh/yr) | 142,940 | 1,158,980 |
| Water Consumption (1000 gal/yr) | 8,860 | 71,800 |
| Capital Cost (Million \$) | \$250.2 | \$1,354 |
| Annual O&M Costs (Million \$/yr) | \$9.9 | \$57.7 |
| By-Product Credit (Million \$/yr) | \$13.0 | \$105.4 |
| Non-feedstock Production Cost (\$/gal) | \$3.66 | \$1.94 |

Tijmensen, Faaij and van Hardeveld (2002) project the cost of FT-liquids is about \$2.70/gal for an 80 tph facility (about 700,000 dry ton per year).⁴⁵ The production of FT-distillates from the model for a similar sized facility using a feedstock cost of \$37.90/dry ton for poplar is \$3.10/gal. Although this result is slightly higher than the cost suggested by Tijmensen et al. (2002), these results cannot be directly compared as the FT-liquids from the model are separated into components. In the conversion technology model, the naptha is sold as a lower value by-product, at a value of about \$1.50/gallon.

3.4 Lignocellulosics to Mixed Alcohols: Thermochemical Conversion

3.4.1 Technology Description

The conversion of LignoCellulosics to Mixed Alcohols via ThermoChemical Synthesis (LCMA Synthesis/TC) is very similar to the process described above for syngas conversion to ethanol (LCE Fermentation/TC), except the fermentation step is replaced by a catalyst conversion.⁴⁶ It is also similar to the LCMD – FT process, although the end products are alcohols not distillates. Figure 11 shows a block diagram for a typical LCMA Synthesis/TC process. Higher Alcohol Synthesis (HAS) is done in reactors similar to those used for FT and methanol synthesis. The most effective types of catalysts include modified methanol synthesis catalysts, modified FT catalysts, and alkali-doped molybdenum catalysts (Nexant Inc. 2006). The modified methanol synthesis and Mo catalysts show higher alcohol yields than modified FT catalysts. The Mo catalysts also have the best selectivity for higher alcohols, and a high tolerance for CO₂ and sulfur in the syngas (Nexant Inc. 2006). However, although this higher sulfur tolerance requires less clean-up of the syngas before conversion, it may require sulfur removal downstream in the mixed alcohol fuel.

⁴⁵ Production cost converted to 2006\$ using CEPCI.

⁴⁶ See section 3.2.1 for further details.

Figure 11 - Schematic diagram of thermochemical ethanol production process

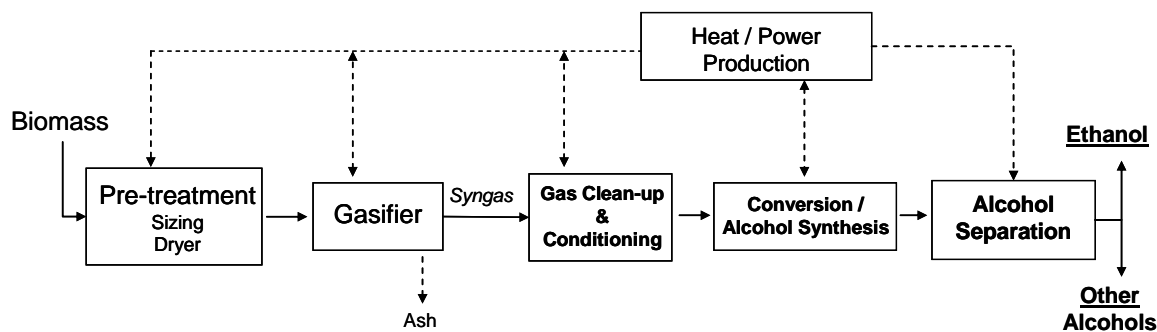


Diagram based on Figure in Phillips et al. (2007)

LCMA Synthesis/TC involves complex set of reactions that produce a variety of products, depending on catalyst used and process conditions. The major reactions include methanol synthesis, FT reactions, higher alcohol synthesis, and water-gas shift. The process is optimized at syngas compositions with ratios of $H_2/CO \sim 1$ (Spath and Dayton 2003). Regardless of the type of catalyst used, typically 40 to 90% of product stream needs to be recycled to maximize mixed alcohol production (Nexant Inc. 2006). The main by-products of the process are CO_2 and water, and large quantities of methane are also often produced.

Technology Status

Although the process to generate higher alcohols from syngas has been known since the early 1900's, the technology has not yet reached commercialization (Spath and Dayton 2003). As of April 2005, there were no commercial plants that solely produce mixed alcohols in the C_2 to C_6 range (Nexant Inc. 2006). However, Range Fuels, Inc. is planning a demonstration facility to generate ethanol and other alcohols via thermochemical conversion (see details below). The main technical hurdles for higher alcohol synthesis include poor selectivity to higher alcohols and low yields. Typical conversion rate for single pass processes is about 10% production of alcohols, which is mostly methanol (Spath and Dayton 2003). The methanol can be recycled back through process to generate higher alcohols.

Research and development efforts for HAS have been performed by several companies since the early 1980's. Some of the most advanced processes were developed by DOW, IFP and Snamprogetti (Spath and Dayton 2003). However, none of these companies are currently active in this area of research (Nexant Inc. 2006). Recent efforts for commercialization of the process have been spurred by new catalyst developments, new project developers, and the interest in alternative fuels (Nexant Inc. 2006). Some of the current commercialization efforts are described below based on information from Nexant Inc. (2006):

- Pearson Technologies has a 30 ton/day biomass gasification with syngas conversion to alcohols in Aberdeen, Mississippi. Pearson is also trying to develop a demonstration plant in Hawaii.
- As mentioned above, Range Fuels Inc. is developing a 1,200 ton/day (wood chip input) demonstration facility to generate ethanol and methanol. This project is supported by DOE and is scheduled for completion in 2011 (see U.S. DOE 2007).
- Power Energy Fuels is continuing to work on their proprietary Ecalene™ process. Although this process is currently only developed at the bench-scale, there are 2 or 3 pilot plants under consideration which would produce mixed alcohols from biomass sources (wood chips, RDF, and tires). One of these pilot facilities is a 2,000 gallon per day plant that would be located at Wabash River Coal Gasification facility (with ConocoPhillips).

- Standard Alcohol Company of America is continuing to work on their Envirolene™ process. This is only a bench scale process currently, but the company is interested in developing a pilot.

Despite this recent surge of development, there are still a number of technical and economic hurdles that must be overcome for the commercialization of the LCMA Synthesis/TC process. According to Nexant Inc. (2006, p. 3-2), the primary technical barriers include: “the overall process feasibility to produce the desired product slate, the ability to scale-up the process to a commercial level, the appropriate process conditions both in the reactor and upstream units, performance of various catalysts at commercial conditions, catalyst sensitivities, and appropriate syngas compositions.” Additionally, large-scale mixed alcohol synthesis will require detailed consideration of the market interest and production costs relative to other fuels.

3.4.2 Performance and cost analysis

The performance and cost model for thermochemical ethanol production via mixed alcohol synthesis has been developed primarily based on recent work by NREL (Phillips et al. 2007). In this report, NREL developed a detail process model and economic analysis for a projected thermochemical ethanol facility. The analysis is based on the DOE targets for synthesis yields and selectivity for a 2,205 dry ton/day facility (equivalent to 772,000 ton/yr for a plant with 96% availability).

The NREL model configuration uses an indirect steam gasifier and a conventional steam power cycle. The syngas clean-up and conditioning step includes tar reforming, water scrubbing (for cooling and quench), and acid gas removal. Clean syngas is converted to alcohols in a fixed bed reactor, using a MoS₂ catalyst with a very high ethanol selectivity.⁴⁷ The alcohol separation section includes dehydration and separation of alcohols. Methanol is recovered and recycled through the alcohol synthesis section to increase yield of ethanol and higher alcohols.

A portion of the unconditioned syngas is diverted to generate electricity and heat. Although this model produces exactly the amount of energy required to sustain the process (consuming 28% of the syngas), an actual plant could vary the energy production depending on favorable market conditions to buy or sell electricity from the grid.

If all the syngas was converted to alcohols, the ethanol yield would be 110.9 gallons per dry ton of biomass, and the total alcohol production would be 130.0 gal/ton (Phillips et al. 2007). With 28% of the syngas used for energy production, the ethanol yield is 80.1 gal/ton.

Outline for model analysis

The key values and assumptions used to calculate the advanced technology analysis and economic performance model for thermochemical ethanol production via higher alcohol synthesis are described below. An example of the detailed analysis is shown in Appendix G. All aspects of the model analysis are based on data from Phillips et al. 2007, unless otherwise noted.

⁴⁷ This is a modified Fischer Tropsch catalyst based on the former Dow/UCC catalyst, with conversion performance modeled based on target results. In addition to higher total CO conversion and higher alcohol selectivity, the projected distribution of ethanol and methanol used in the NREL model differ from current results. The Phillips et al. (2007) model assumes that 71% of the alcohol production is ethanol, and 5% is methanol. In contrast, the Dow distribution was 30-70% methanol and 34.5% ethanol.

- Feedstock type and input quantity are variables. Applicable feedstocks include woody and agricultural residues. The moisture content of these feedstocks must be less than 50% for optimal operation.
- This model is applicable for facility sizes with greater than 185,000 dry ton biomass input per year, following the size range for Fischer Tropsch synthesis which has similar technology characteristics.
- As different feedstocks have distinct heating values, the feedstock choice affects the syngas produced in the gasifier and thus the alcohol conversion. Yield of ethanol and other alcohols are calculated based on the HHV conversion efficiency. Ethanol conversion is equivalent to 38.9% of the input energy from biomass, converted to gallons using a typical ethanol HHV of 84,000 Btu/gal. The yield of other alcohols is based on an HHV efficiency of 7.6%. Nearly all of this stream consists of propanol, which has an HHV of about 97,300 Btu/gal.
- Annual water consumption is estimated to be 1.94 gallons per gallon of ethanol produced. This water use includes cooling tower make-up, process water, and other system losses.
- Capital cost data is based on values from Phillips et al. 2007 for a 2,205 dry ton/day plant and a scaling factor of 0.8 which is typical for biofuel conversion facilities.
- Annual Fixed O&M costs include a labor costs and other fixed O&M. Labor costs are calculated following Phillips et al. 2007, with a base cost of \$4,720,000/yr for a 772,000 ton/yr facility. The labor costs are scaled by facility size using a scaling factor of 0.25, following the Fischer Tropsch analysis (see above for details). The other fixed O&M costs (maintenance, insurance, and taxes) are calculated as a percentage (4%) of the total capital investment.
- Annual Variable O&M costs include consumables (such as catalysts, gasifier bed material, and other raw materials) and waste disposal. These are calculated on a feedstock input basis. No electricity purchase is required, as a portion of the syngas is diverted for heat and power production.
- The co-product credit for mixed alcohols is calculated following the methodology of Phillips et al. 2007. The value of the alcohols is estimated to be \$1.4/gal, based on the current wholesale gasoline prices from EIA of \$1.95/gal (average from January 2006 through April 2007), a 78% discount for lower energy density, and a 10 ¢/gal reduction for potential additional processing requirements before use.⁴⁸
- The levelized non-feedstock production cost for the ethanol is calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results projected for LCMA Synthesis/TC are shown in Table 14. This table shows results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs. The low end feedstock input shows the minimum applicable facility size, and the high end was selected to show a reasonable range of sizes and costs. In general, the analysis shows that the non-feedstock production cost of ethanol from thermochemical conversion of biomass to mixed alcohols is relatively low, although it is slightly higher than the non-feedstock production cost of ethanol from the LCE Fermentation/Hydrolysis process.

⁴⁸ The 10 ¢/gal discount is applied to account for the fact that no special effort is made to ready the mixed alcohols for blending or sale.

Table 14 - Example Cost and Performance for Ethanol from LCMA Synthesis/TC

| Sample Model Results - LCE Fermentation/TC w/ Poplar | | |
|---|---------------|---------------|
| Feedstock Input (dry ton/yr) | 185,000 | 1,500,000 |
| Ethanol Yield (MGY) | 15.4 | 125 |
| Other Alcohols Yield (MGY) | 2.6 | 21 |
| Water Consumption (1000 gal/yr) | 29,920 | 242,570 |
| Capital Cost (Million \$) | \$64.9 | \$346 |
| Annual O&M Costs (Million \$/yr) | \$9.7 | \$20.3 |
| By-Product Credit (Million \$/yr) | \$3.6 | \$29.5 |
| Non-feedstock Production Cost (\$/gal) | \$1.38 | \$0.74 |

Phillips et al. (2007) estimate the minimum selling price of thermochemical ethanol is \$1.08/gal for a 772,000 dry ton per year facility.⁴⁹ The ethanol production cost from the model for a similar sized facility using a delivered feedstock cost of \$35/dry ton for poplar is \$1.260/gal. As the Phillips et al. (2007) study is the basis of this analysis, it is likely that different economic factors and analysis methods contribute to this variation.

The LCMA Synthesis/TC process is very similar to LCMD – FT, such that one would expect similar capital cost projections and biofuel yields for both processes. However, a comparison of the example results from the technology characterization models (Table 13 and Table 14) portrays LCMA Synthesis/TC as having much lower capital costs and higher yields than the FT process. It is possible that this is due to more optimistic technology development projections for LCMA Synthesis/TC. There do not seem to be any published studies showing experimental support for the targets given in Phillips et al. 2007. As such, the LCMA Synthesis/TC technology will not be included in the Western supply curve model. However, it is important to point out that this technology uses similar feedstocks and has a similar production cost as LCE Fermentation/Hydrolysis, so this should not have a large effect on the overall supply curve.

3.5 Lignocellulosics to Gasoline: Upgrading Pyrolysis Oil

LignoCellulosics to Gasoline via Pyrolysis and Upgrading (LCG Upgrading/Pyrolysis) is the conversion of biomass to bio-oil via fast pyrolysis, and then upgrading via hydrotreatment followed by hydrocracking at a petroleum refinery to produce a hydrocarbon fuel. In this process the bio-oil is mixed with petroleum products during some stage in the refining process, which enables usage of advanced refinery technologies and economies of scale. In the following discussion the general method of biomass pyrolysis will first be reviewed, followed by an analysis of the development of bio-oil co-processing in refineries.

3.5.1 Technology Description

Biomass Pyrolysis

Pyrolysis of biomass is a rapid thermal process conducted in an environment without oxygen to prevent combustion. By controlling temperatures and reaction times, pyrolysis can be used to convert solid biomass materials into a liquid oil comprised of low molecular weight fragments of the lignin, cellulose and hemicellulose. This “bio-oil” can be used directly as a substitute for petroleum fuel oil in boilers and gas turbines. Alternately, bio-oil can be further treated to produce refined fuels or natural chemicals such as adhesives, resins,

⁴⁹ Production cost converted from 2005\$ to 2006\$ using CEPCI.

polymers and flavorings. A variety of biomass feedstocks have been used for pyrolysis, including wood, bark, paper, bagasse, corn fibers and other agricultural residues. The bio-oil yield primarily depends on the processing conditions with the rate of heat transfer and product vapor residence time being the key parameters. Feedstock composition, principally the ash component, can also be a factor but plays a lesser role. Under optimum process conditions liquid yields of 60-80% by weight can be realized (Ensyn Group Inc. 2001).

To prepare the biomass for fast pyrolysis, the feedstock is dried to less than 10% moisture content and sized to small relatively homogeneous particles. This is important to ensure a rapid heat transfer. In commercial pyrolysis processes, the reaction occurs in a fluidized bed reactor using an inert material such as sand to transfer heat to the incoming biomass particles. This process occurs so fast that the biomass is flash vaporized, becoming a mixture of gas, vapor, aerosols and solid char. Fast pyrolysis produces higher yields of the liquid fraction in bio-oil than conventional pyrolysis, with a typical composition of 75% liquid, 12% char, and 13% gas (Ringer, Putsche and Scahill 2006).⁵⁰

The pyrolysis step typically occurs at temperatures around 450-500°C, at atmospheric pressure, with a reaction time from less than one second to several seconds. After the char is separated out (using a cyclone), the vapor product is rapidly cooled and condenses into liquid bio-oil. The remaining non-condensable gas is recycled within the process and used as a fluidizing gas for pyrolysis and as a fuel combined with the solid char to provide process heat.

The bio-oil product is an oxygenated fuel that consists primarily of carbon, hydrogen and oxygen and small amount of nitrogen and sulfur. The exact composition of the product depends on the feedstock composition and the reaction conditions.

Status of commercialization

There are currently two main companies that have developed commercial pyrolysis oil technologies - Ensyn Corporation and DynaMotive Energy Systems Corporation. Each company has a patented pyrolysis technique and a wide range of experience producing bio-oil from different feedstocks. Their products have been tested and used in various applications. By 2005 Ensyn had 7 commercial RTP™ biomass plants, and the largest plant can process 160 green tons of wood per day (Ensyn Corporation n.d.). DynaMotive Energy Systems Corporation reached commercialization in 2004 with a 110 ton/day biomass pyrolysis BioTherm™ plant in Ontario. DynaMotive also completed their first modular 220 ton/day biomass pyrolysis plant in December 2006 (Dynamotive Energy Systems Corporation 2006).

Ringer, Putsche and Scahill (2006) note that although pyrolysis oil production is technically commercialized, most of these commercial facilities are fairly small and the larger ones primarily produce products such as flavoring compounds which have limited markets. Only a small amount of bio-oil is currently produced for fuel.

Co-processing bio-oil in petroleum refineries

Crude pyrolysis oil can be used directly to produce heat and power. However, upgrading bio-oil to a hydrocarbon fuel via hydrotreatment and hydrocracking can significantly increase the value of the product by producing fungible gasoline and diesel fuels. Hydrotreating and hydrocracking are well-developed processes currently used for rejecting

⁵⁰ Conventional pyrolysis involves lower reaction temperatures and long residence times, and typically results in nearly equal fractions of liquid, gas, and char produced (Bridgwater 2007).

nitrogen, oxygen and other heteroatoms⁵¹ from crude petroleum oils. This potential upgrading approach is the focus of efforts to co-refine pyrolysis oils with petroleum.

One of the main issues with crude bio-oil is the lack of stability over time. According to Ringer, Putsche and Scahill (2006), bio-oil viscosity increases with time at much faster rate than petroleum products. Such highly viscous oil cannot be used as motor fuel. Furthermore, the phase change which accompanies the increase in viscosity is also a problem for usage. Char fines in the bio-oil seem to be a critical factor causing this lack of stability. Advanced technologies that can successfully remove fine particles can significantly increase the shelf life of the product and are currently under development.⁵²

Other potential issues with bio-oil include re-vaporization from liquid state and potential environmental and health effects. Bio-oil has a very complex chemical composition, and more than 300 specific compounds have identified so far (Ringer, Putsche and Scahill 2006). Some of the compounds present are known carcinogens, and experimental tests have shown mixed results with respect to health risks. Further, although bio-oil biodegrades in soil more easily than petroleum-based hydrocarbon fuels, its acidity can be harmful to water environments in the event of large spills (Ringer, Putsche and Scahill 2006).

Additionally, incorporation of new fuels into a mature and accepted framework can be difficult. Using bio-oil directly as a transportation fuel would likely require engine modifications and new infrastructure. However, "if bio-oils could be upgraded chemically to produce a product that looked more like petroleum hydrocarbons then the end use device would require little to no modification. This would be the fastest way to gain acceptance of biomass-based fuels into the existing infrastructure." (Ringer, Putsche and Scahill 2006, p. 27)

The bio-oil upgrading process primarily involves oxygen removal by hydrotreatment followed by hydrocracking. Catalytic hydrotreating and cracking deoxygenate the bio-oil and reform the remaining carbon and hydrogen into hydrocarbon compounds in the gasoline and diesel boiling point range. Since the carbon oxides and water are removed, these processes reduce the yield of the liquid product of bio-oil (by weight), although hydrotreating gives higher yields than catalytic cracking (Ringer, Putsche and Scahill 2006). The hydrocarbon product from catalytic cracking processes is also dominated by aromatic compounds which limits the amount that can be blended into motor fuels (Ringer, Putsche and Scahill 2006).

Pyrolysis oil contains a mixture of products including water-soluble oxygenated compounds (derived from the hemicellulose), and insoluble pyrolytic lignin (UOP 2005). The pyrolytic lignin is a lower molecular weight version of lignin, which results from the thermal conversion of biomass feedstocks. This component of the pyrolysis oil is lower in oxygen and has a higher energy content than the water soluble portion. It can be separated out via gravity separation or by adding water to the pyrolysis oil to precipitate out the insoluble pyrolytic lignin portion. Hydrotreatment of the pyrolytic lignin requires less hydrogen than the pyrolysis oil because of the lower oxygen content, and can also be done at mild reaction conditions.

⁵¹ A heteroatom is an organic chemistry term to describe any atom that is not carbon or hydrogen (Source: Wikipedia).

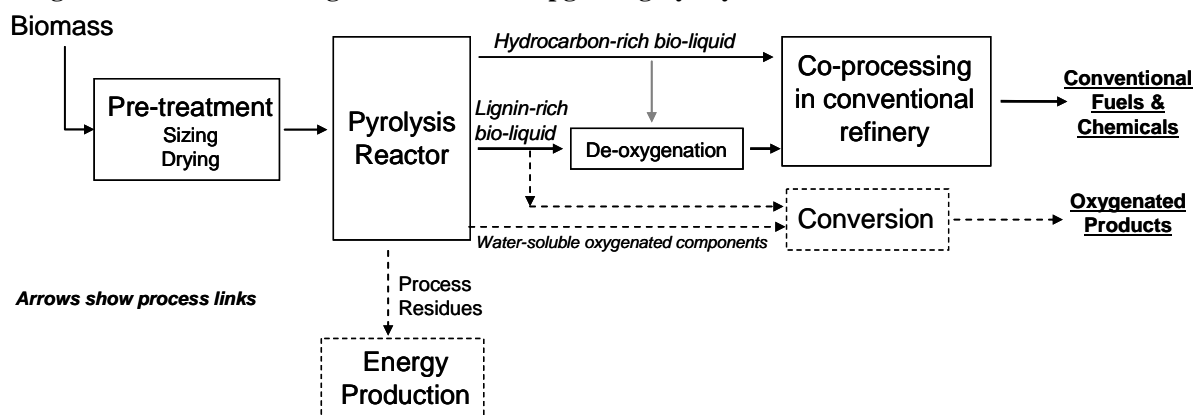
⁵² A 6-month shelf (equivalent to that of petroleum products) life can be achieved with the addition of alcohol or other solvents and removing the char fines. The fine particle removal is difficult and somewhat analogous to gas-clean up from gasification. Additional research is needed in order to find an easy, effective, low-cost solution.

High oxygen and acid content is a challenge for processing bio-oils and other biomass products. Typical refinery feedstocks like crude oil have little or no oxygen, but pyrolysis oils have oxygen contents on the order of 35-45% (UOP 2005). The pyrolytic lignin has less oxygen than the water-soluble carbohydrate portion of the bio-oil, and therefore requires less hydrogen consumption during the upgrading process. However, the carbohydrate portion of bio-oil has been shown to be a promising feedstock for reforming to hydrogen which could provide a synergy with co-processing of both components of the bio-oil. The water soluble portion can also be used as a fuel to generate heat and power for the upgrading process.

Pyrolysis oils are much more acidic than other refined fuels. The acidity of refinery processed feedstocks is measured by total acid neutralization (TAN) number.⁵³ The TAN for crude is <1, while a typical pyrolysis oil has TAN of 78 (UOP 2005). This level of acidity requires processing in stainless steel vessels, which are expensive and not standard in refineries. Pretreatment of the pyrolysis oils can be done in stainless steel vessels to reduce the acidity, making co-processing in the downstream vessels possible.

Figure 12 shows a potential arrangement for producing pyrolysis oil and co-processing in a petroleum refinery. The biomass feedstock is first sized and dried for pyrolysis. After the reaction, the pyrolysis oil is separated into components - the oil fraction and pyrolytic lignin are co-processed to produce conventional fuels while the water-soluble components are converted into oxygenated products. Process residues are used for energy generation.

Figure 12 – Schematic Diagram of the LCG Upgrading/Pyrolysis Process



Based on figure in (Solantausta 2006)

One of the key barriers for development of the co-processing technology is that refineries are unwilling to experiment with processing for small quantities of pyrolysis oil that may detrimentally affect their catalysts or end product (UOP 2005). A typical U.S. petroleum refinery generates 150,000 barrels per day (bpd), while the total production of pyrolysis oil is only a small fraction of that amount (UOP 2005).

BIOCOUP is a consortium of research partners that have formed to evaluate the development of co-processing bio-liquids in refineries. According to a presentation given at the 2006 European Conference on Biorefinery Research (Solantausta 2006), the consortium's objectives for pyrolysis oil co-processing include: reducing bio-oil production

⁵³ Total acid neutralization number is defined as the number of milligrams of potassium hydroxide to neutralize on gram of feedstock (UOP 2005).

costs, develop and scale de-oxygenation technology for upgrading bio-oil, assess the viability of co-processing upgraded bio-liquids, optimize recovery and fractionation strategies for conversion to chemicals, and to develop life cycle analysis and scenario analysis.

3.5.2 Performance and cost analysis

Bio-oil yield is affected by a number of factors, including feedstock composition, reaction temperature and heat transfer rate, and amount of inert gas in the reaction environment. Furthermore, the mineral matter in biomass acts as a catalyst for cracking and polymerization during pyrolysis, which affects the composition of the bio-oil product (Ringer, Putsche and Scahill 2006). However, despite the importance of many factors in bio-oil production, the heating value of the product is relatively constant (on a dry basis) at about 7,500-8,000 Btu/lb (Ringer, Putsche and Scahill 2006). This is only a slight reduction from the heating values of the biomass feedstocks used for pyrolysis, which are generally around 8,500 Btu/lb.

The water content in the biomass feedstock is also an important factor in bio-oil production. In fact, even pyrolysis of bone-dry biomass produces bio-oil with 12-15 wt% water (Ringer, Putsche and Scahill 2006). Any moisture in the input biomass will contribute to this percentage. As water is an unfavorable component on bio-oil, which also acts as a heat sink, it is advantageous to remove as much water as possible from the incoming biomass before pyrolysis.

Upgrading bio-oil is important for a number of reasons. Hydrotreating and hydrocracking pyrolysis oil converts it to a hydrocarbon fuel that can be blended and transported with petroleum fuels. This provides access to a wide infrastructure already in place. Upgrading can also significantly increase the value of the bio-oil.

The ratio of pyrolytic lignin and water soluble pyrolysis oils is such that in the near term, a substantial electricity/steam plant should accompany development of any fuel project based on this technology to ensure maximum use of the incoming resource. As complete and efficient use of incoming feedstock is critical to any project, the energy/liquid fuel product ratio will be an important consideration in siting. A potential alternative for the long-term is to reform the water-soluble portion of the bio-oil to generate hydrogen for use in the upgrading process. Although the reforming technology is currently too expensive to support a stand-alone pyrolysis oil upgrading facility, it is a possible future development.

The evolution of pyrolysis oil production and upgrading is assumed to be as follows:

- *Short term*: no pyrolysis oil upgrading technologies available (still in research phase)
- *Mid term*: hydrotreating / hydrocracking of pyrolytic lignin and oil fraction in separate (skid-mounted) process co-located at a refinery; resulting products blended with petroleum products for storage and distribution.
- *Long term*: no analysis of the technology for this period. However, hydrotreating / hydrocracking the pyrolytic lignin and oil fraction via co-processing in a stand-alone refinery, and reforming the water-soluble portion of bio-oil for hydrogen production is a potential long-term scenario.

Outline for model analysis

The key values and assumptions used to calculate the mid-term technology analysis and economic performance model for LCG Upgrading/Pyrolysis are described below. An example of the detailed analysis is shown in Appendix G.

- Feedstock type and quantity are variables for input. Types of materials that can be used for pyrolysis include woody and agricultural feedstocks. The applicable size range for the model is biomass input of 30,000 - 800,000 dry tons per year, based on reported facility sizes in UOP 2005; Ringer, Putsche and Scahill 2006; and Polagye, Hodgson and Malte 2007.
- The facility is assumed to be operational at full load 330 days per year, which is equivalent to a capacity factor of 90%.
- Pyrolysis yield depends on feedstock type, and the values used are based on published data from Ensyn and Dynamotive (Ensyn Group Inc. 2001; Dynamotive Energy Systems Corporation 2000).
- The conversion efficiency of biomass to pyrolysis oil is estimated to be 66-83% (on an energy basis) depending on the feedstock, with bark giving the lowest yield and bagasse and mixed paper giving the highest. This is calculated using the HHV of the feedstocks and typical pyrolysis oil properties.
- Pyrolytic lignin (including the oil fraction) is the only portion of the bio-oil that undergoes hydrotreatment / hydrocracking for upgrading to hydrocarbon fuels, and the hydrotreatment and hydrocracking units are sized based on this feed stream.
- The pyrolytic lignin content (weight percentage) for all herbaceous feedstocks based on values for bagasse, and percentage for woody feedstocks is based on data for softwood (from Dynamotive Energy Systems Corporation 2000).
- The only significant by-products of pyrolysis oil production are solid char and non-condensable gases. It is assumed both of these products are recycled into the process to use as a heat source and fluidizing gas for the reaction. There are several products from upgrading, including gasoline and diesel, light hydrocarbons and the un-treated water-soluble bio oil. Gasoline is the primary product, accounting for 30% of the converted pyrolytic lignin (by weight).
- The diesel and light hydrocarbon by-products are valued at \$1.76/gal and \$0.93/gal respectively, based on the average 2005 wholesale values from EIA AER 2005.⁵⁴
- The water-soluble bio-oil can be cofired with natural gas (or refinery off-gas) to provide heat and/or electricity. This material needs to be co-fired because of its relatively high water content (20-30%) and low heating value (5,000 Btu/lb on a moisture-free basis). If the water-soluble bio-oil was not combusted it would become a disposal liability. To simplify the model, it is assumed that the value of the heat/electricity produced from co-firing the water-soluble bio-oil offsets the cost of natural gas consumed for combustion. As such, there is no net value or cost associated with the water-soluble component of the bio-oil.
- The yield of the concentrated CO₂ generated from hydrotreatment is derived from UOP 2005. No value is assigned to this stream in the analysis.
- Water consumption is estimated based on data from Ringer et al. 2006. According to Ringer et al. (2006), a 200,750 dry ton per year facility uses about 160 thousand gal/hr water, mostly for cooling. It is assumed that only 2% of this water is lost in blowdown, so that the annual consumption is about 28,000 thousand gallons per year. This value is scaled based on feedstock input. No additional water consumption is added for hydrotreatment.
- Capital costs for the pyrolysis process and hydrotreatment are calculated separately, as they have different size and economies of scale.
- Total capital investment values for pyrolysis oil production for a range of sizes were available from various sources.⁵⁵ These values were used to determine a relationship

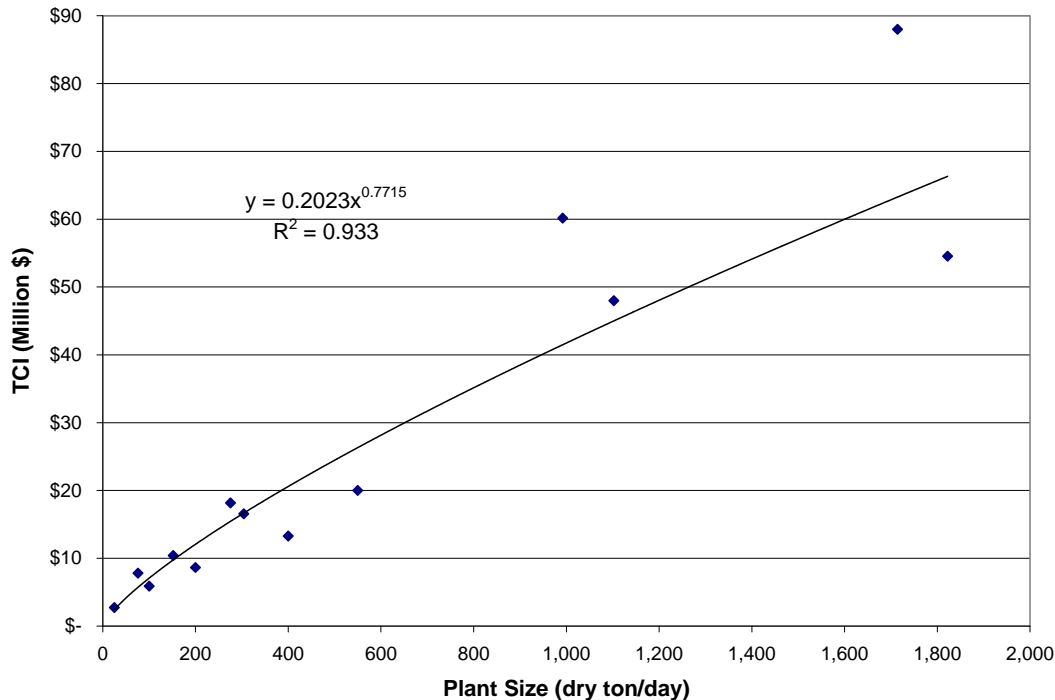
⁵⁴ For comparison, the projected values of diesel and Liquefied Petroleum Gas in 2015 are \$2.04/gal and \$1.34/gal (\$633/ton), respectively (EIA AEO 2007).

⁵⁵ Sources: Ringer, Putsche and Scahill 2006; Polague et al. 2007; UOP 2005; and Cole Hill Associates 2004.

between facility size and capital cost, shown in Figure 13. A scaling factor of 0.8 is used for capital cost calculations.

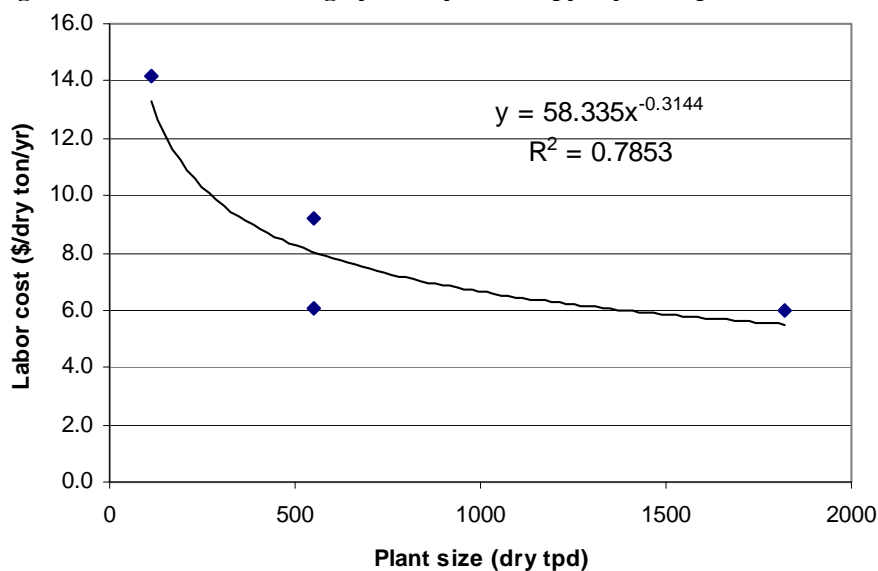
- The capital cost for hydrotreatment is based on a value of \$30 million for a unit that processes 2,250 barrels per day, and a standard scaling factor of 0.6 (UOP 2005).
- Variable O&M costs for the process include electricity and other utilities, hydrogen gas for hydrotreatment, and waste disposal. These annual costs are estimated individually based on plant size. Electricity cost for pyrolysis oil production has been updated to 5.7 cents/kWh based on the 2005 average cost for electricity in the industrial sector from EIA (2006). The electricity cost for hydrotreatment has not been updated as there was not sufficient detail in the published studies, but the difference is expected be minimal and would not have significant effect on the production cost.
- Annual fixed O&M costs include labor, maintenance, insurance, overhead, and other costs. Except for the labor component, these costs are estimated as a percentage of total capital cost. The labor portion only applies to the pyrolysis oil production process, and is scaled based on project size using a relation derived from the available data from Polagye et al. (2007) and Ringer, Putsche and Scahill (2006). This relationship is shown in Figure 14.
- The levelized non-feedstock production costs for gasoline are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Figure 13 – Capital cost and plant size relation for fast pyrolysis



Based on data from Cole Hill Associates 2004, Ringer et al. 2006, Polagye et al. 2007, UOP 2005.

Figure 14 – Labor cost scaling by facility size for pyrolysis oil production



Based on data from Polagye et al. 2007; and Ringer, Putsche and Scahill 2006.

Model Results

Some of the key cost and performance results projected for LCG Upgrading/Pyrolysis are shown in Table 15. This table shows results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs. In general, the analysis shows that the non-feedstock production cost of gasoline with this method is quite expensive. This suggests that further development is needed to make pyrolysis oil with upgrading competitive with other biofuel technologies. For example, co-processing in a refinery could have significant impact on the production cost if the processing challenges can be overcome.

Table 15 - Example Cost and Performance for LCG Upgrading/Pyrolysis

| Sample Model Results - LCG Upgrading/Pyrolysis w/ Wood | | |
|--|---------------|---------------|
| Feedstock Input (dry ton/yr) | 30,000 | 800,000 |
| Crude Pyrolysis Oil Yield (MGY) | 4.6 | 122 |
| Gasoline Yield (MGY) | 0.5 | 13.5 |
| Diesel Yield (MGY) | 0.1 | 3.3 |
| Light Hydrocarbon Yield (ton/yr) | 839 | 22,380 |
| CO ₂ Stream (ton/yr) | 7,600 | 202,620 |
| Water Consumption (1000 gal/yr) | 4,180 | 111,420 |
| Capital Cost - pyrolysis (Million \$) | \$6.6 | \$82.6 |
| Capital Cost - hydrotreatment (Million \$) | \$4.1 | \$29.4 |
| Annual O&M Costs (Million \$/yr) | \$2.0 | \$35.1 |
| By-Product Credit (Million \$/yr) | \$0.6 | \$15.7 |
| Non-feedstock Production Cost (\$/gal) | \$5.32 | \$2.45 |

3.6 Fatty Acids to Hydrocarbon (Renewable Diesel): Hydrotreatment

Green diesel and renewable diesel are some of the terms used to describe a hydrocarbon fuel produced from biomass oil feedstocks to differentiate them from biodiesel generated via

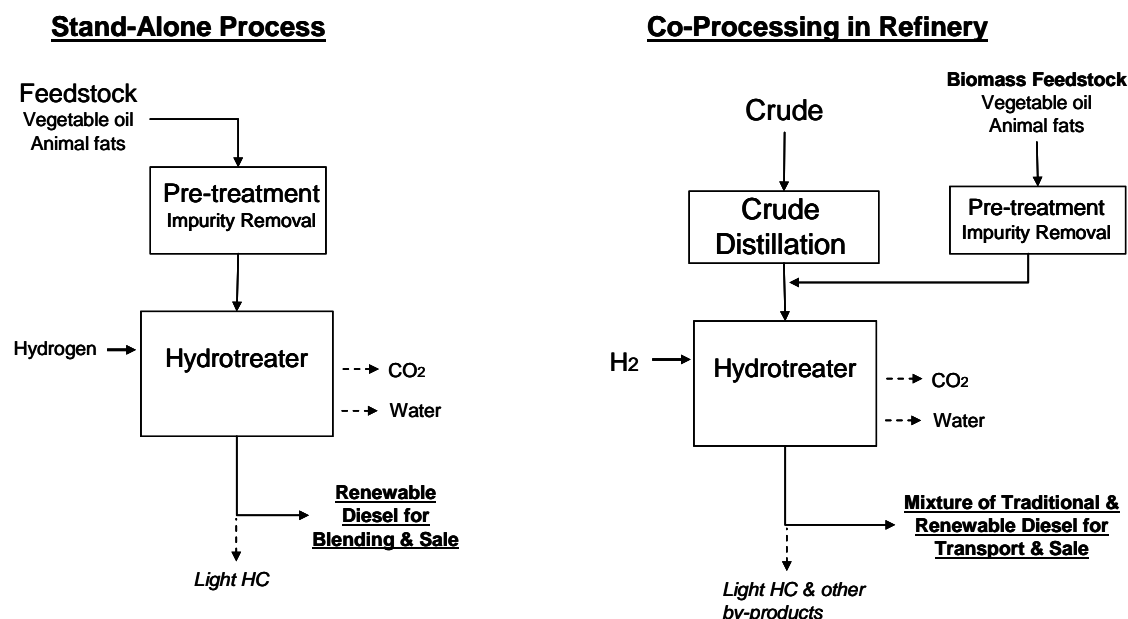
transesterification. Although renewable diesel is produced from vegetable oils and waste greases, unlike fatty acid methyl ester (FAME) biodiesel this product does not have oxygen in its molecular structure. It can also utilize oil feedstocks with higher amounts of free fatty acids than the FAME transesterification process. Furthermore, renewable diesel has superior properties to petroleum-based diesel and can be blended readily as it is a hydrocarbon fuel (UOP 2005). For example green diesel typically has less than 10 ppm Sulfur, has cetane numbers as high as 100, and cloud point temperatures as low as -30° C (which minimizes cold weather gel issues). And unlike with FAME production, there is no glycerin co-product produced, which could become a waste disposal issue for that industry. Fatty Acids to HydroCarbon (FAHC) renewable diesel also has a superior emission profile when compared to either conventional petroleum based diesel or FAME biodiesel. Particulates, NO_x, HC, and CO are all reduced whereas biodiesel has sometimes shown higher levels of NO_x (Climate Change 2006). Renewable diesel can be produced in petroleum refineries to take advantage of economies of scale and similar process requirements, which helps minimize the cost of production.

3.6.1 Technology Description

The Fatty Acids to HydroCarbon – Hydrotreatment (FAHC – Hydrotreatment) production process for renewable diesel in petroleum refineries is very similar to the process for upgrading pyrolysis oils described above, except that it does not typically involve hydrocracking. However, the Super Cetane Technology developed by the CANMET Energy Technology Center (CETC) at Natural Resources Canada does involve a hydrocracking step to tailor the molecular weight of some of the hydrocarbon compounds. They have reported cetane numbers exceeding 100 with their technology (Natural Resources Canada n.d.). As with the hydrotreatment / hydrocracking of pyrolysis oil, upgrading biomass oil feedstocks can be done in a stand alone skid-mounted unit or pretreated and co-processed with petroleum feedstocks. Schematic diagrams of these configurations are shown in Figure 15.

In the FAHC - Hydrotreatment process, vegetable oil or animal fat is upgraded using hydrogen gas, producing diesel and propane or other light hydrocarbons. Some water and CO₂ are also released in the process. A heterogeneous catalyst is used for hydrotreatment, which tend to be less expensive than homogeneous catalyst required for FAME biodiesel (UOP 2005). The product distribution is typically 83-86% diesel and 2-5% light hydrocarbons by weight (in terms of incoming biomass) (UOP 2005). By comparison, the FAME process results in approximately 96% diesel production. This higher biodiesel yield from FAME is due to the retention of oxygen in the product. In hydrotreatment the oxygen is rejected in the form of CO₂ or H₂O.

Figure 15 – Schematic diagram of renewable diesel production in petroleum refinery



Oils and greases typically have an oxygen content of 11-12%, which is much lower than pyrolysis oils but still significantly higher than petroleum (UOP 2005). As with pyrolysis oil hydrotreatment / hydrocracking, the acidity of the oils generally means that they must be processed in stainless steel vessels. This is particularly true for waste greases which tend to have higher acidity than virgin feedstocks (UOP 2005).

UOP analyzed co-processing in a distillate hydrotreater for ultra-low sulfur diesel, and found that for brown grease feedstocks "standard refinery distillate hydrotreating units do not appear to be suitable for green diesel production in a co-processing scheme." (UOP 2005, p. 18). In addition to the potential affect on metallurgy requirement, co-processing can detrimentally affect catalyst lifetime. This is partly due to the fact that hydroprocessing conditions for standard hydrotreating for ultra-low sulfur diesel in refinery favors hydrodeoxygenation (HDO), which produces water from biomass feedstocks and can have a negative affect on catalyst activity. UOP (2005) further noted that although co-processing with brown grease had a positive effect on catalyst in terms of sulfur, higher heat release causes increased deactivation rate which lowered the catalyst lifetime.

Stand alone, skid-mounted processing of biomass oil and grease feedstocks seems to be an effective alternative, as it avoids some of the issues caused by co-processing but still has some benefits from refinery integration. In this configuration, biomass feedstocks are first pre-processed to remove contaminants, typically using some combination of the following: hydroclones, dewatering, acid washing, ion exchange, desalting, fixed guard bed catalyst. The feedstocks can then be hydrotreated with a skid mounted system. A benefit of skid-mounted system is the ability to optimize the reaction for biomass feedstocks, so that the hydrotreatment conditions favor decarboxylation (DeCO_2) instead of HDO. The DeCO_2 process removes oxygen from the feedstock as CO_2 instead of water, which can be captured relatively easy. DeCO_2 also reduces hydrogen consumption, as it only requires 0.8 wt% H_2 whereas HDO requires 3%.

Both skid-mounting and co-processing are potential treatment options in refineries. Despite the current processing issues, it is likely that with further technology developments and

pretreatment methods, co-processing of waste greases could be an economically favorable process in the long-term. Furthermore, several of the companies developing renewable diesel refining use co-processing techniques with vegetable oils and animal fats, which suggests this process is already favorable for virgin feedstocks.

Status of Commercialization

ConocoPhillips has completed a commercial demonstration of FAHC renewable diesel production at Whitegate refinery in Cork, Ireland, which co-processes vegetable oil with crude oil. The facility was able to produce 42,000 gallons per day of renewable diesel product (Green Car Congress 2006). ConocoPhillips has also partnered with Tyson to develop a facility to process animal fats in the US. This facility is planned for start up in late 2007, and will be able to produce up to 500,000 gallons per day of renewable diesel (ConocoPhillips 2007).

Neste also has developed a process to generate renewable diesel, using stand-alone hydrotreatment of vegetable oils and animal fats. Their NExBTL (NExT Biomass To Liquid) process will be demonstrated at the Porvoo oil refinery in Finland, with a 60 MGY renewable diesel facility (Neste Oil n.d.). In late 2006 Neste Oil announced plans to build a second 60 MGY plant adjacent to the first plant to meet growing domestic demand. In addition Neste Oil has another FAHC renewable diesel production facility under development with OMV in Austria. Although the NExBTL process is stand-alone, it will take advantage of existing refinery infrastructure for energy, blending, logistics, and laboratories (Koskinen, Nurminen and Sourander 2005). For example, a refinery-based NExBTL plant has readily available access to hydrogen and utilities. Additionally, the renewable diesel can be blended, stored, and transported with petroleum based diesel which improves process economics and logistics.⁵⁶

The Petrobras H-BIO process uses co-processing of vegetable oils to generate FAHC renewable diesel. Petrobras has tested a range of oils and process conditions, and the process has a conversion rate of 95% by volume or higher (Petrobras n.d.). Petrobras is planning to have H-BIO process operations in at least 3 refineries by the end of 2007, with a total capacity to process more than 250,000 tons of vegetable oil annually. Further implementation in 2 more refineries is planned for 2008 (Petrobras n.d.).

Others companies that are currently involved in FAHC renewable diesel production include Nippon Oil in Japan which expects to have a commercial process within 3 years, and BP which has a 80,000 gallon per day demonstration facility planned for start-up in Australia this year (ConocoPhillips 2007).

3.6.2 Performance and cost analysis

Outline for model analysis

The key values and assumptions used to calculate the technology analysis and economic performance model for stand-alone and co-processing FAHC - Hydrotreatment processes are described below. An example of the detailed analysis is shown in Appendix G.

- Feedstock input quantity is a variable.
- Applicable feedstocks are virgin oil, animal fats, and yellow grease. However, feedstock choice does not affect the product yield in this model. The typical heating values of the

⁵⁶ NExBTL diesel is compatible with existing fuel logistics and vehicles, and meets diesel fuel standards in all properties except density, which is about 780 kg/m³ (Neste Oil n.d.). As such, the density value is the only blending constraint and limits blends of NExBTL product to 65% by volume (Koskinen, Nurminen and Sourander 2005).

feedstocks are very similar, generally about 16,500 to 17,000 Btu/lb HHV (US EPA 2001). Although processing of yellow grease feedstocks may lead to higher capital costs as a result of additional pre-processing (such as desalting) and different metallurgy requirements due to higher acidity, the potential added cost has not been included as it is expected to be minimal (on the order of a few percent according to discussions with industry contacts).

- The applicable size range for the model is 15 to 200 MGY of renewable diesel, based on current and planned facility sizes for ConocoPhillips, Neste, and Petrobras. The smallest facility reported by these companies can produce around 1,000 barrels per day of renewable diesel, and the largest (planned) facility will be able to generate about 12,000 barrels per day.
- Stand-alone FAHC renewable diesel yield is 86 wt% based on data from UOP (2005). Co-processing yield is 81 wt% based on data for Petrobras H-BIO system (Petrobras n.d.). The renewable diesel yields are converted to volumetric quantities using a typical density of 6.5 lb/gal.
- The conversion efficiency of oil/fat to fuel (on an energy basis) is estimated to be 98% for skid-mounted and 96% for co-processing. This is calculated using an average LHV of oil feedstocks of 16,000 Btu/lb, and a typical renewable diesel (lower) heating value of 123,200 Btu/gal.
- Hydrogen consumption is calculated as a percentage of the feedstock input (by weight). The stand-alone system favors DeCO₂, which requires less hydrogen than the HDO reaction favored in co-processing. According to UOP (2005), typical hydrogen consumption ranges from 1.5 - 3.8 wt% of feedstock input. The low end value is used for the stand-alone hydrotreatment in the model, while the high end value is used for co-processing.
- Light hydrocarbon production is estimated to be 3.5 wt% for stand alone process and 4.4 wt% for co-processing, based on input feedstock (UOP 2005, Petrobras n.d.)
- Water and CO₂ are also generated during hydrotreatment. The processing conditions for the stand alone process favor CO₂ production, while conditions for co-processing typically produce water. The total amount of CO₂ and water produced is calculated as the difference between the inputs and other outputs, which give results similar to UOP (2005).⁵⁷
- Water consumption data is not included for hydrotreatment processes, as it is assumed to be minimal.
- Near term capital costs are based on reported costs of facilities under development from Neste (stand alone) and Petrobras (co-processing), and a scaling factor of 0.6.⁵⁸
- Mid term capital costs are decreased based on next generation facility development and a learning rate of 20% (McDonald and Schrattenholzer 2002). This rate is similar to that found for ethanol production facilities and retail petrol processing.
- Variable O&M costs include utilities and hydrogen. The utility cost is based on data for pyrolysis oil hydrotreatment from UOP (2005). Annual hydrogen cost is based on hydrogen consumption and cost data from UOP 2005.
- Fixed O&M costs include maintenance, insurance, and overhead. These costs are estimated to be 5.5% of near term capital cost, based on pyrolysis oil hydrotreatment O&M. No labor cost is added, as renewable diesel production is part of larger refinery operation and will not require additional operators.
- The propane/light hydrocarbon co-product value is calculated using the average 2005 wholesale price of \$0.93/gal (EIA AER 2005).⁵⁹ The estimated added cost for compressing propane is negligible and has not been included.⁶⁰

⁵⁷ UOP (2005) estimates that the total water and CO₂ output is equivalent to 12-16 wt% of feedstock input.

⁵⁸ This scaling factor is equivalent to the value used for pyrolysis hydrotreatment, following UOP (2005).

⁵⁹ For comparison, the projected value of Liquefied Petroleum Gas is projected to be \$1.34/gal (\$633/ton) in 2015 (EIA AEO 2007).

- The levelized non-feedstock production costs for FAHC - Hydrotreatment are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results projected for the FAHC - Hydrotreatment processes are shown in Table 16 and Table 17. These tables show results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs. In general, the analysis shows that the non-feedstock production cost of FAHC renewable diesel via co-processing in a refinery is cheaper than a stand-alone process, as expected. However, it seems that either method could be used to produce biofuel economically. The projected non-feedstock costs on a \$/gallon basis are more than competitive with FAME biodiesel, and the resulting fuel is higher quality and can be readily integrated into existing infrastructure.

Table 16 - Example Cost and Performance for FAHC - Hydrotreatment Stand Alone Process

| Sample Model Results - FAHC Hydrotreatment (skid-mounted) - mid term | | |
|---|---------------|---------------|
| Feedstock Input (ton/yr) | 20,000 | 785,000 |
| Renewable Diesel Yield (MGY) | 5.1 | 200 |
| Light Hydrocarbon Yield (ton/yr) | 700 | 27,475 |
| Capital Cost (Million \$) | \$25.8 | \$233.3 |
| Annual O&M Costs (Million \$/yr) | \$2.2 | \$31.1 |
| By-Product Credit (Million \$/yr) | \$0.3 | \$12.1 |
| Non-feedstock Production Cost (\$/gal) | \$0.98 | \$0.24 |

Table 17 - Example Cost and Performance for FAHC - Hydrotreatment Co-Processing

| Sample Model Results - FAHC Hydrotreatment (co-processing) - mid term | | |
|--|---------------|---------------|
| Feedstock Input (ton/yr) | 20,000 | 800,000 |
| Renewable Diesel Yield (MGY) | 5.0 | 200 |
| Light Hydrocarbon Yield (ton/yr) | 880 | 35,200 |
| Capital Cost (Million \$) | \$4.1 | \$37.3 |
| Annual O&M Costs (Million \$/yr) | \$1.3 | \$41.5 |
| By-Product Credit (Million \$/yr) | \$0.4 | \$15.5 |
| Non-feedstock Production Cost (\$/gal) | \$0.27 | \$0.15 |

For perspective, the projected 2015 price for crude soybean oil is about \$681.40/ton (FAPRI 2007). Including an estimated delivery cost of \$10/ton,⁶¹ the feedstock cost for renewable diesel production is \$2.71/gal for a skid-mounted process and \$2.77/gal for co-processing. This gives a total production cost of \$2.95 to \$3.69 per gallon for stand-alone and \$2.92 to \$3.04 per gallon for co-processing.⁶² These costs are very similar to the projected production costs for FAME biodiesel from virgin oil in 2015, which range from \$2.98 to \$3.36 per gallon for facilities from 1 to 80 MGY (see Section 2.2.2 for details). For comparison, the national average wholesale price for diesel over the past year is \$2.03 per gallon (EIA July 2007). Also note that the current retail biodiesel price is about \$3.20/gal (USDA AMS 2007).

⁶⁰ Antares estimates that compressing 17,000 tons per year of propane would cost estimate about \$100K/yr, using an electricity price of 9¢/kWh.

⁶¹ Based on a truck transportation cost of \$0.20/ton-mile and a 50 mile delivery radius.

⁶² The production cost does not include distribution costs or profit, which are both factored into retail fuel prices.

4 Emissions Data for Biofuel Conversion Technologies

Water use, greenhouse gas and criteria pollutant emissions data were estimated for each modeled conversion technology, in order to be used as inputs for the environmental benefits analysis in Task 5. These data were calculated based on the conversion technology models or available data in the literature. Additional information about data sources and calculations are included below.

4.1 Yields and Water Use

Table 18 summarizes the water use and fuel yields for various conversion technologies. The biomass conversion technology data are based on the technology models developed for this effort. The crude oil refining data is included for comparison, based on results from Sheehan et al. 1998.

Table 18 – Water Use and Yield for Various Conversion Technologies

| Technology | Feedstock | Yield (gal/ton)* | Water Use (gal / gal product) |
|--|-----------------------------|------------------|-------------------------------|
| Biomass Conversion Technologies | | | |
| Grain to Ethanol (Dry Mill) | Grains | 100.0 | 4.7 |
| Grain to Ethanol (Wet Mill) | Grains | 89.3 | 24.4 |
| FAME Biodiesel | Virgin Oil | 258.2 | 0.03 |
| | Waste Oil | 249.1 | 0.03 |
| | Animal Fat | 266.3 | 0.03 |
| LCE Fermentation / Enzymatic Hydrolysis (dilute acid pretreatment) | Wood Chips (hardwood) | 85.9 | 6.2 |
| | Wood Chips (softwood) | 90.2 | 5.9 |
| | Corn Stover / Ag Residues | 80.6 | 6.6 |
| | Wheat Straw / Ag Residues | 76.8 | 7.0 |
| | Switchgrass / HEC | 77.4 | 6.9 |
| LCMD - Fischer Tropsch | Wood Chips (hardwood) | 40.6 | 1.2 |
| | Wood Chips (softwood) | 42.0 | 1.1 |
| | Corn Stover / Ag Residues | 36.8 | 1.3 |
| | Wheat Straw / Ag Residues | 38.7 | 1.2 |
| | Switchgrass / HEC | 37.7 | 1.3 |
| LCG - Upgrading/Pyrolysis | Clean Wood Chips | 16.9 | 8.2 |
| | Softwood | 15.7 | 8.9 |
| | Bark | 14.2 | 9.8 |
| FAHC - Hydrotreatment (stand-alone) (co-processing) | Vegetable Oil or Animal Fat | 255.4 | 0.0 |
| | Vegetable Oil or Animal Fat | 250.5 | 0.0 |
| Crude Oil Refining | | | |
| Gasoline Refining* | Petroleum | -- | 0.003 |
| Diesel Refining* | Petroleum | -- | 0.001 |

* Yield reported in gallons per dry ton except for Grains (in Grain to Ethanol), which are in as rec'd tons (15% MC)

** Data from Sheehan et al. 1998

4.2 Greenhouse Gas and Criteria Pollutant Emissions

Table 19 shows the reported emissions resulting from biofuel conversion, based on data from various published Life Cycle Analysis (LCA) studies. These studies use detailed analyses of all the significant energy use and emissions impacts from the entire biofuel production process, including emissions released in the conversion process, emissions associated with energy use (electricity and steam), and emissions allocated to the production of consumables used in the process. Publicly available LCA data for biofuel conversion processes is limited – the only processes with published results are the current conversion technologies (grain to ethanol and FAME biodiesel from virgin oil), and LCE via fermentation and enzymatic hydrolysis for some feedstocks. Although there are some LCA studies available for other advanced biofuel conversion processes, they typically only report the overall results of a well-to-wheels analysis. These are insufficient for our analysis, since several aspects of the analysis are location-dependant, including feedstock production. As such, an analysis of the CO₂ emissions has been developed based on the conversion technology models and data from other various sources. This will be discussed below in detail.

Table 19 - Summary of Conversion Technology Emissions from Published Life Cycle Analyses

| Technology | Feedstock | Total GHG (g CO _{2-eq} /gal)* | GHG Emissions (g/gal) | | | Criteria Pollutants (g/gal) | | | | Allocation basis | Source |
|--|----------------------------|---|-----------------------|-----------------|------------------|-----------------------------|------|-----------------|-----------------|----------------------------------|---------------------|
| | | | CO ₂ | CH ₄ | N ₂ O | CO | PM | SO ₂ | NO _x | | |
| Biomass Conversion Technologies | | | | | | | | | | | |
| Grain to Ethanol (Dry Mill) | Grains | 2,032.1 | n/a | n/a | n/a | n/a | n/a | n/a | n/a | mass | Kim and Dale 2005 |
| Grain to Ethanol (Wet Mill) | Grains | 2,376.3 | n/a | n/a | n/a | n/a | n/a | n/a | n/a | mass | Kim and Dale 2005 |
| FAME Biodiesel | Soybean Oil | 606.0 | 566.5 | 1.67 | 0.00 | 0.21 | 0.60 | 8.17 | 1.36 | mass | Sheehan et al. 1998 |
| LCE Fermentation / Enzymatic hydrolysis (dilute acid pretreatment) | Wood (50%), Grass (50%) | n/a | n/a | 0.21 | n/a | 5.67 | 2.35 | 0.87 | 3.52 | credit for displaced electricity | Delucchi 2003 |
| | Corn Stover | 8,921.5 | 9,051.1 | n/a | n/a | n/a | n/a | n/a | n/a | credit for displaced electricity | Sheehan et al. 2004 |
| Crude Oil Refining | | | | | | | | | | | |
| Gasoline Refining | Petroleum | 1,671.2 | 1,530.6 | 5.65 | 0.04 | 1.55 | 0.44 | 1.56 | 3.11 | n/a | Delucchi 2003 |
| Diesel Refining | Petroleum | 909.1 | 837.5 | 2.21 | 0.07 | 1.30 | 0.31 | 1.83 | 1.56 | n/a | Delucchi 2003 |

* For cases in which the Total GHG Emissions was not provided, this value was calculated using 100-year Global Warming Potentials from IPCC 2001 to convert the GHG emissions into CO_{2-eq} factors (CO₂ (1); CH₄ (23); N₂O (296)).

Note that the data shown in Table 19 only correspond to the portion of the emissions attributed to the conversion of the feedstock to fuel; it does not include feedstock production, end use of the fuel, or any other component of the life cycle analysis.

Several conversion technologies produce byproducts in addition to biofuels (such as glycerin in the FAME Biodiesel process), so the environmental impacts must be distributed among the products. An allocation method is used to determine the portion of the emissions that are attributable to biofuels for each process. There are several methods for allocation,

including by product weight, energy content, value, or replacement products.⁶³ Each method will give different results for the environmental impacts of biofuels. Although some authors argue that the replacement method is the most robust (Kim and Dale 2005), the mass and energy allocation methods are frequently used due to simplicity and ease of application. All of the LCA studies reported here use the mass based allocation method.

Table 20 shows the calculated CO₂ emissions from biofuel conversion technologies based on estimations from the characterization models and other sources. Definitions, data sources and assumptions are given below. Note that these emissions do not include those related to the production and use of chemical consumables in the processes. The LCA study from Sheehan et al. 1998 showed that the effect of the consumables can make up a large portion of the overall GHG emissions. For example, the GHG emissions associated with production of chemicals used in the FAME biodiesel process makes up more than 40% of the total CO₂ emissions. As such, the total CO₂ emissions for FAME Biodiesel with virgin oil shown in Table 20 are much smaller than those in Table 19. However, the energy usage portion of the emissions from Sheehan et al. 1998 (364 g CO₂-eq/gal) are very similar to the emissions estimated from this analysis (344 g CO₂/gal).

Definitions:

- The **total Gross CO₂** values includes emissions related to grid electricity use, steam/thermal energy use, conversion process, on-site hydrogen production, and stack and other CO₂ emissions for the entire conversion process.
- The **biofuel Gross CO₂** is the biomass portion of the total gross CO₂, using the mass basis allocation shown in the table.
- The **avoided CO₂** from net electricity production is the emissions that are avoided by displacement of grid electricity in processes that produce power as a byproduct.
- The **Biofuel Net CO₂** is the resulting emissions attributed to biomass, less the avoided CO₂ from electricity production.
- The **Total CO₂ from Biomass** is reported to differentiate from the fossil fuel CO₂. The biomass portion of the CO₂ emissions may be excluded from the total, as biomass can be considered net neutral for carbon due to its uptake in growth cycles. The biomass CO₂ includes any emissions from the conversion process itself, and CO₂ emitted from the stack during biomass fueled cogeneration.

⁶³ For the replacement products method, emissions associated with materials that would be used to replace the product are tallied and subtracted from the gross process emissions. For example, for distiller's dried grains, the emissions due to production of an alternate animal feed such as soybean meal would be used.

Table 20 – CO₂ emissions (g/gallon) from biofuel conversion based on technology model data

| Technology | Feedstock | Allocation (mass basis) | Gross CO ₂ (total) | Gross CO ₂ (biofuel) | Avoided CO ₂ (electricity) | Net CO ₂ (biofuel) | Total CO ₂ from Biomass |
|--|------------------------------|----------------------------|----------------------------------|------------------------------------|---|----------------------------------|--|
| <i>Relationships Equations</i> | | <i>A</i> | <i>B</i> | <i>C = A x B</i> | <i>D</i> | <i>E = C – D</i> | <i>F</i> |
| Grain to Ethanol (Dry Mill) | Grains | 50% | 6,096 | 3,024 | 0 | 3,024 | 2,835 |
| Grain to Ethanol (Wet Mill) | Grains | 51% | 8,426 | 4,281 | 0 | 4,281 | 2,835 |
| FAME Biodiesel | Soybean Oil | 81% | 425 | 344 | 0 | 344 | 0 |
| | Waste Oil | 81% | 1,214 | 984 | 0 | 984 | 0 |
| | Animal Fat | 81% | 425 | 344 | 0 | 344 | 0 |
| LCE Fermentation / Enzymatic Hydrolysis (dilute acid pretreatment) | Wood Chips (hardwood) | 100% | 11,217 | 11,217 | 1,228 | 9,989 | 11,217 |
| | Wood Chips (softwood) | 100% | 11,199 | 11,199 | 1,228 | 9,971 | 11,199 |
| | Corn Stover / Ag Residues | 100% | 9,663 | 9,663 | 1,228 | 8,434 | 9,663 |
| | Wheat Straw / Ag Residues | 100% | 9,578 | 9,578 | 1,228 | 8,349 | 9,578 |
| | Switchgrass / HEC | 100% | 9,756 | 9,756 | 1,228 | 8,527 | 9,756 |
| LCMD - Fischer Tropsch | Wood Chips (hardwood) | 75% | 25,728 | 19,399 | 11,612 | 7,787 | 25,728 |
| | Wood Chips (softwood) | 75% | 27,761 | 20,932 | 11,612 | 9,320 | 27,761 |
| | Corn Stover / Ag Residues | 75% | 25,559 | 19,272 | 11,612 | 7,660 | 25,559 |
| | Wheat Straw / Ag Residues | 75% | 23,198 | 17,491 | 11,612 | 5,879 | 23,198 |
| | Switchgrass / HEC | 75% | 28,201 | 21,264 | 11,612 | 9,652 | 28,201 |
| LCG - Upgrading/Pyrolysis (1) | Clean Wood Chips | 55% | 91,941 | 50,291 | 0 | 50,291 | 79,584 |
| | Softwood | 55% | 108,762 | 59,493 | 0 | 59,493 | 95,874 |
| | Bark | 55% | 128,890 | 70,503 | 0 | 70,503 | 115,250 |
| FAHC – Hydrotreatment (2) (stand alone) (co-processing) | Soybean Oil | 96% | 2,756 | 2,646 | 0 | 2,646 | 531 |
| | Soybean Oil | 95% | 3,684 | 3,496 | 0 | 3,496 | 653 |

1). No thermal energy credits or consumption are included here as sufficient details on the process are not available. Although some of the by-products are used for energy production (such as char from pyrolysis, and water-soluble bio-oil for upgrading), fossil fuels may be needed to help combust some of these materials (such as the water-soluble bio-oil which has a 30% MC).

2). One of the output streams from this process is a mixture of water and CO₂. Although the ratios of each component depend on the process conditions, here we assume equal amounts of each.

Note that emissions for LCG – Upgrading/Pyrolysis are much higher than any other conversion technology, as a result of the low yields from the upgrading process.

Data Sources and Assumptions:

- Grid electricity use was determined using conversion technology models where possible. Emissions associated with grid electricity use based on the Western Region average from EIA 2002.
- Electricity production from LCE Fermentation / Enzymatic Hydrolysis and LCMD – Fischer Tropsch based on conversion technology characterization models. Avoided emissions from net electricity exported to grid are based on the Western Region average from EIA 2002. Total process energy for LCE Fermentation / Enzymatic Hydrolysis based on data from Hamelinck et al. 2005.
- Emissions associated with fossil fuel derived thermal energy are based on emissions factors from Sheehan et al. 1998 for steam production in an industrial boiler using coal and natural gas.
- Data for energy use for dry mill Grain to Ethanol process from Shapouri and Gallagher 2005; data for wet mill process from Shapouri et al. July 2002. Kim and Dale 2005 suggest the thermal energy for wet milling is about 20% natural gas and 80% coal. It is assumed that all of the thermal energy production from dry milling comes from natural gas.
- Energy requirements for FAME Biodiesel for base catalyzed process (virgin oil and animal fats) are based on data from Sheehan et al. 1998 and Haas et al. 2006. According to Zhang et al. 2003, the energy requirements for the acid catalyzed process (waste grease) are 3 times higher and the electricity requirements are twice as high as base catalyzed process. For both FAME Biodiesel processes it is estimated that all thermal energy is produced from natural gas.
- Thermal energy use for FAHC – Hydrotreatment based on data for diesel hydrotreatment in a refinery from Wang et al. 2004.
- Hydrogen use for LCG – Upgrading/Pyrolysis and FAHC – Hydrotreatment are from technology characterization models. Emissions from on-site hydrogen production based on natural gas refining from Contadini et al. 2000.
- Stack and other CO₂ emissions were determined by via difference in carbon between the biomass input and product outputs.
- Conversion process CO₂ emissions are from the conversion technology models.

References

- Aden, A, et al. *Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis for Corn Stover*. NREL technical report no. NREL/TP-510-32438. 2002.
- Antares Group Inc. *Biomass Integrated Gasification - Fischer Tropsch (BIG-FT), Phase 2 / Report 1*. Prepared for US DOE Biomass Program and NREL. 2003.
- . *Biomass Integrated Gasification - Fischer Tropsch (BIG-FT), Phase 2 / Report 2*. Prepared for US DOE Biomass Program and NREL. 2003.
- . *Biomass Integrated Gasification - Fischer Tropsch (BIG-FT), Phase 2 / Report 3*. Prepared for US DOE Biomass Program and NREL. 2003.
- . *Quantifying Biomass Resources for Hydrothermal Processing*. Prepared for Battelle Memorial Institute. August 17 2005.
- . *Quantifying Biomass Resources for Hydrothermal Processing II*. Prepared for Battelle Memorial Institute. November 30 2005.
- Bender, Martin. "Economic Feasibility Review for Community-Scale Farmer Cooperatives for Biodiesel." *Bioresource Technology* 70 (1999): 81-87.
- Bioengineering Resources, Inc. *Technology Summary*. February 7 2007
<<http://www.brienergy.com/pages/process01.html>>.
- Bridgwater, Tony. *Biomass Pyrolysis*. IEA Bioenergy Task 34: Pyrolysis of Biomass. 2007.
- Brown, Robert C., and Jennifer Holmgren. "Fast Pyrolysis and Bio-Oil Upgrading."
- Butzen, Steve, and Troy Hobbs. "Corn Processing III: Wet Milling." *Pioneer High-Bred International Crop Insights* 12.15 (2002).
- Canakci, Mastafa, and John Van Gerpen. "A Pilot Plant to Produce Biodiesel from High Free Fatty Acid Feedstocks." Paper presented at the ASAE Annual International Meeting, 2001.
- Cole Hill Associates. *Bio-Oil Commercialization Plan*. Prepared for the NH Office of Energy and Planning. 2004.
- ConcoPhillips. "Renewable Diesel." Briefing Presentation, 2007.
- . *Renewable Diesel Briefing (Presentation)*. 2007.
- Contadini, J. Fernando, et al. "Hydrogen Production Plants: Emissions and Thermal Efficiency Analysis." Paper presented at the Second International Symposium on Technological and Environmental Topics in Transports. Milan, Italy, October 21, 2000.
- Climate Change. *NExBTL Renewable Synthetic Diesel*. 2006
<http://www.climatechange.ca.gov/events/2006-06-27+28_symposium/presentations/CalHodge_handout_NESTE_OIL.PDF>.
- Delucchi, Mark A. *A Lifecycle Emissions Model (LEM): Lifecycle Emissions from Transportation Fuels, Motor Vehicles, Transportation Modes, Electricity Use, Heating and Cooking Fuels, and Materials*. Institute of Transportation Studies, UC Davis no. UCD-ITS-RR-03-17. 2003.
- Dynamotive Energy Systems Corporation. *200 TPD BioOil Plant Takes Shape in Guelph, Ontario*. January 18 2006
<<http://www.dynamotive.com/english/news/releases/2006/december/061218.html>>
- . *Fast Pyrolysis of Biomass for Green Power Generation*. Presentation at the first world conference and exhibition on biomass and energy for industry. 2000.
- EIA. *Annual Energy Outlook 2007 with Projections to 2030*. Tech. Rept. No. DOE/EIA-0383(2007). 2007.
- . *Annual Energy Review 2005*. Tech. Rept. No. DOE/EIA-0384(2005). 2005.
- . *Average Retail Price of Electricity to Ultimate Customers by End-Use Sector*. October 4 2006.

- . *No. 2 Distillate Prices by Sales Type*. July 26 2007
<http://tonto.eia.doe.gov/dnav/pet/pet_pri_dist_dcu_nus_m.htm>.
- . "Updates State-Level Greenhouse Gas Emissions Coefficients for Electricity Generation 1998-2000," 2002.
- Energy Research Centre of the Netherlands. *Phyllis, Database for Biomass and Waste*. n.d.
<<http://www.ecn.nl/phyllis>>.
- Ensyn Corporation. *Who We Are*. n.d. April 3 2007
<<http://www.ensyn.com/who/ensyn.htm>>.
- Ensyn Group Inc. *The Conversion of Wood and Other Biomass to Bio-Oil*. 2001.
- EUBIA. *Biofuels for Transport*. 2006 <<http://www.eubia.org/>>.
- FAPRI (Food and Agricultural Policy Research Institute). *FAPRI 2007 U.S. and World Agricultural Outlook*. FAPRI Staff Report no. 07-FSR 1. 2007.
- FAPRI (Food and Agricultural Policy Research Institute). *FAPRI 2008 U.S. and World Agricultural Outlook*. FAPRI Staff Report no. 08-FSR 1. 2007.
- Gallagher, Paul W., Heather Brubaker, and Hosein Shapouri. "Plant Size: Capital Cost Relationships in the Dry Mill Ethanol Industry." *Biomass and Bioenergy* 28 (2005): 565-71.
- Green Car Congress. *ConocoPhillips Begins Production of Renewable Diesel Fuel at Whitegate Refinery*. January 20 2006.
- Haas, Michael J, et al. "A Process Model to Estimate Biodiesel Production Costs." *Bioresource Technology* 97 (2006): 671-78.
- Hamelinck, Carlo N, et al. *Production of FT Transportation Fuels from Biomass; Technical Options, Process Analysis and Optimisation, and Development Potential*. Utrecht University. 2003.
- Hamelinck, Carlo N., Geertje van Hooijdonk, and Andre PC Faaij. "Ethanol from Lignocellulosic Biomass: Techno-Economic Performance in Short-, Middle- and Long-Term." *Biomass and Bioenergy* 28 (2005): 384-410.
- Industrial Information Resources. *Ethanol and Biodiesel Database*. 2007
<<http://www.industrialinfo.com/index.jsp>>.
- Kim, Seungdo, and Bruce E. Dale. "Environmental Aspects of Ethanol Derived from No-Tilled Corn Grain: Nonrenewable Energy Consumption and Greenhouse Gas Emissions." *Biomass and Bioenergy* 28 (2005): 475-89.
- Koskinen, Matti, Matti Nurminen, and Mauri Sourander. "The Comprehensive Biofuel Strategy for a Refinery." Paper presented at the ERTC 10th Annual Meeting, November 16, 2005.
- Lynd, L. R, et al. *Strategic Biorefinery Analysis: Analysis of Biorefineries*. NREL subcontractor report no. NREL/SR-510-35578. 2005.
- Ma, Fangrui, and Milford A. Hanna. "Biodiesel Production: A Review." *Bioresource Technology* 70 (1999): 1-15.
- McAloon, Andrew, et al. *Determining the Cost of Producing Ethanol from Corn Starch and Lignocellulosic Feedstocks*. NREL technical report no. NREL/TP-580-28893. 2000.
- McDonald, Alan, and Leo Schrattenholzer. "Learning Curves and Technology Assessment." *International Journal of Technology Management* 23.7/8 (2002): 718-45.
- McKeever, David B., and Kenneth E. Skog. Urban Tree and Woody Yard Residues. USDA Forest Service Forest Products Laboratory no. Research Note FPL-RN-0290. 2003.
- Mohan, Dinesh, Charles U. Pittman, and Philip H. Steele. "Pyrolysis of Wood/Biomass for Bio-Oil: A Critical Review." *Energy & Fuels* 20 (2006): 848-89.
- Moilanen, Antero. *Thermogravimetric Characterisations of Biomass and Waste for Gasification Processes*. VTT publications no. 607, ed. Maini Manninen. 2006.
- Natural Resources Canada. *The CETC Supercetane™ Technology*. n.d.
<http://www.rncan.gc.ca/es/etb/ctec/cetc01/htmldocs/Publications/factsheet_supercetane_technology_e.htm>.

NBB. *Biodiesel Production*. n.d.
 <http://www.biodiesel.org/pdf_files/fuelsheets/Production.PDF>.

Neste Oil. *Biodiesel*. n.d. May 24, 2007
 <<http://www.nesteoil.com/default.asp?path=1,41,535,547,3716,3884>>.

Nexant Inc. *Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup, and Oxygen Separation Equipment. Task 9: Mixed Alcohols from Syngas - State of Technology*. NREL subcontractor report no. NREL/SR-510-39947. 2006.

Nilles, Dave. "Combating the Glycerin Glut." *Biodiesel Magazine*. September 2006.
 <http://www.biodieselmagazine.com/article.jsp?article_id=1123&q=glycerin&category_id=26>.

Petrobras. *H-Bio Process*. n.d. May 8 2007
 <<http://www2.petrobras.com.br/tecnologia/ing/hbio.asp>>.

Phillips, S., et al. *Thermochemical Ethanol Via Indirect Gasification and Mixed Alcohol Synthesis of Lignocellulosic Biomass*. NREL technical report no. NREL/TP-510-41168. 2007.

Plastemart. *Feedstock Trends*. 2007. May 21, 2007
 <<http://www.plastemart.com/feedstock.asp>>.

Polagye, Brian L., Kevin T. Hodgson, and Philip C. Malte. "An Economic Analysis of Bio-Energy Options Using Thinnings from Overstocked Forests." *Biomass and Bioenergy* 31 (2007): 105-25.

RFA. *How Ethanol is Made*. 2005 <<http://www.ethanolrfa.org/resource/made/>>.

Ringer, M., V. Putsche, and J. Scahill. *Large-Scale Pyrolysis Oil Production: A Technology Assessment and Economic Analysis*. NREL technical report no. NREL/TP-510-37779. 2006.

Shapouri, Hosein, and Paul Gallagher, U.S. Department of Agriculture, Office of the Chief Economist, Office of Energy Policy and New Uses. *USDA's 2002 Ethanol Cost of Production Survey*. Agricultural economic report no. 841. 2005.

Shapouri, Hosein, Paul Gallagher, and Michael S. Graboski. *USDA's 1998 Ethanol Cost of Production Survey*. Agricultural economic report no. 808. 2002.

Sheehan, John, et al. *Life Cycle Inventory of Biodiesel and Petroleum Diesel for Use in an Urban Bus*. Prepared by NREL for U.S. DOE Office of Fuels Development and U.S. Department of Agriculture's Office of Energy. 1998.

Sheehan, John, et al. "Energy and Environmental Aspects of Using Corn Stover for Fuel Ethanol." *Journal of Industrial Ecology* 7.3-4 (2004): 117-46.

Solantausta, Yrjo. *Co-Processing of Upgraded Bio-Liquids in Standard Refinery Units*. October 19 2006. European Conference on Biorefinery Research
 <http://ec.europa.eu/research/energy/pdf/gp/gp_events/biorefinery/bs3_01_solantausta_en.pdf>.

Solomon, Barry D., Justin R. Barnes, and Kathleen E. Halvorsen. "Grain and Cellulosic Ethanol: History, Economics, and Energy Policy." *Biomass and Bioenergy* (2007).

Spath, Pamela L., and David C. Dayton. *Syngas Analysis - Preliminary Screening, Technical Briefs, and Technical Barrier Assessment for Syngas to Fuels and Chemicals*. National Renewable Energy Laboratory. June 30 2003.

Tijmensen, Michael J. A., Andre P. C. Faaij, Hamelinck, and Martijn R. M van Hardeveld. "Exploration of the Possibilities for Production of Fischer Tropsch Liquids and Power Via Biomass Gasification." *Biomass and Bioenergy* 23 (2002): 129-52.

Tyson, Shaine K, et al. *Biomass Oil Analysis: Research Needs and Recommendations*. NREL technical report no. NREL/TP-510-34796. 2004.

U.S. DOE. *DOE Selects Six Cellulosic Ethanol Plants for Up to \$385 Million in Federal Funding*. February 28 2007 <<http://www.energy.gov/news/4827.htm>>.

- U.S. DOE EERE Biomass Program. *Biomass Feedstock Composition and Property Database*. January 25 2006
<http://www1.eere.energy.gov/biomass/feedstock_databases.html>.
- . *Theoretical Ethanol Yield Calculator*. January 20 2006
<http://www1.eere.energy.gov/biomass/ethanol_yield_calculator.html>.
- U.S. Environmental Protection Agency. *Biodiesel Emissions Database*. 2001
<<http://www.epa.gov/otaq/models/biodsl.htm>>.
- UOP. *Opportunities for Biorenewables in Oil Refineries, Final Technical Report*. Submitted to U.S. Department of Energy. 2005.
- USDA. "Corn Belt Feedstuffs." *USDA-MO Dept Ag Market News*. May 22 2007
<http://www.ams.usda.gov/mnreports/sj_gr225.txt>.
- USDA AMS. *Livestock and Grain Market News - Feedstuff Reports*. 2007. June 25, 2007
<<http://www.ams.usda.gov/lsmnpubs/Feedstuff.htm>>.
- USDA NASS. *Crops and Plants*. June 11 2007
<http://www.nass.usda.gov/QuickStats/indexbysubject.jsp?Pass_group=Crops+%26+Plants>.
- Vicente, Gemma, Mercedes Martinez, and Jose Aracil. "Optimisation of Integrated Biodiesel Production. Part I. A Study of the Biodiesel Purity and Yield." *Bioresource Technology* 98 (2007): 1724-33.
- Wang, Michael, Hanjie Lee, and John Molburg. "Allocation of Energy Use in Petroleum Refineries to Petroleum Products." *International Journal of Life Cycle Assessment* 9.1 (2004): 33-44.
- Whims, John with Agricultural Marketing Resource Center, Kansas State University, Department of Agricultural Economics. "Corn Based Ethanol Costs and Margins Attachment 1." 2002. May 23, 2007
<<http://www.agmrc.org/NR/rdonlyres/AA5DBE03-C649-4679-8A29-BCA18E376F2D/0/ksueth1.pdf>>.
- Wooley, Robert, et al. *Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis Current and Futuristic Scenarios*. NREL technical report no. TP-580-26157. 1999.
- Zhang, Y, et al. "Biodiesel Production from Waste Cooking Oil: 1. Process Design and Technical Assessment." *Bioresource Technology* 89 (2003): 1-16.
- . "Biodiesel Production from Waste Cooking Oil: 2. Economic Assessment and Sensitivity Analysis." *Bioresource Technology* 90 (2003): 229-40.

Appendix A

Conversion Technology Status Matrix

| Reference Name | Feedstock | Conversion Technology | Fuel Type | Benefits | Issues/Challenges | Key Players ¹ | Status |
|---|-------------------------------------|---|------------------------------|--|---|--|--|
| Current Representative Technologies | | | | | | | |
| Grain Ethanol - Dry Mill | Grains / Starches | Enzymatic Fermentation | Ethanol | Commercial technology - many facilities already operational and in construction. | Feedstock availability and production is limited, and there is competition with food products. Net energy balance lower than projections for cellulosic ethanol. Limited opportunity for improvement. | POET, Dupont, Abengoa, VeraSun, and others | Commercial |
| Grain Ethanol - Wet Mill | Grains / Starches | Separation and Fermentation | Ethanol | Commercial technology - many facilities already operational and in construction. | | ADM, Cargill, Bunge, Pacific Ethanol, and others | Commercial |
| Fatty Acid Methyl Ester (FAME) | Seed Oil / Waste Oils / Animal Fats | Esterification | Methyl Esters | Ease of conversion, biodiesel can be used in diesel engines without adjustments. Significant heating oil market. | Relatively small scale. Limited opportunity for process improvements. Cold weather technical issues. | ADM, Cargill, Bunge, BlueSun Biodiesel, and others | Commercial |
| Sugar to Ethanol Fermentation | Sugars | Fermentation | Ethanol | Commercial technology. More efficient than conversion of starches. | Limited feedstock availability in North America | Imperial Ethanol Inc., Imperial Bioresources LLC | Commercial |
| Additional Technologies Projected to be in Use by 2015 to 2025 | | | | | | | |
| LignoCellulosics to Ethanol - Fermentation/Hydrolysis (LCE Fermentation/Hydrolysis) | Lignocellulosic Biomass | Enzymatic Hydrolysis and Fermentation | Ethanol | Relatively low maintenance costs. Potentially very high glucose yield. | Cellulase costs must be reduced. Enzymes are sensitive to poisoning. | POET, Abengoa, Iogen, Catalyst, Mascoma (NREL) | Demonstration/pilot plants currently operating. More planned for operation within 5 years. |
| | | Acid Hydrolysis and Fermentation | Ethanol | Technically mature - dilute acid process is oldest cellulosic ethanol technology. | Expensive vessels and high maintenance costs (corrosion). Large amount of gypsum for disposal. Requires improved acid recovery (especially strong acid process). | strong acid - Arkenol, BlueFire Ethanol Inc. dilute acid - (Univ. Wisconsin, NREL) | Demonstration plants are planned for operation within 5 years. |
| LignoCellulosics to Ethanol - Fermentation/ThermoChemical (LCE Fermentation/TC) | Lignocellulosic Biomass | Gasification and Fermentation | Ethanol | Higher yield of ethanol per ton of feedstock than direct fermentation of biomass. See Note #2. | See Note #3. | ALICO, BCT (Univ of Arkansas) | Pilot / demonstration facilities planned for operation within 5 years. |
| LignoCellulosics to Middle Distillates Fischer Tropsch (LCMD - FT) | Lignocellulosic Biomass | Gasification and Fischer Tropsch Synthesis | Middle Distillates, Gasoline | FT diesel can be substituted directly for conventional diesel with lower emissions. See Note #2. | See Note #3. Catalysts sensitive to poisoning and sintering. Requires improved yields. | Bechtel, ECN, Shell, Choren | Pilot facilities in operation, demonstration facilities planned for 2009 and beyond |
| LignoCellulosic to Mixed Alcohol - ThermoChemical (LCMA - TC) | Lignocellulosic Biomass | Gasification and Thermochemical Conversion | Mixed Alcohols | Mixed alcohols are easily blended with gasoline. See Note #2. | See Note #3. Catalysts sensitive to deactivation from sintering. Process requires improved yields. Potential groundwater contamination from mixed alcohols usage. | Range Fuels, Pearson (PNNL, NREL) | Pilot / demonstration facilities planned for operation within 5 years. |
| LignoCellulosics to Gasoline - Upgrading/Pyrolysis (LCG - Upgrading/Pyrolysis) | Lignocellulosic Biomass | Pyrolysis Oil Production and Upgrading via hydrotreatment / hydrocracking | Bio-oil, Diesel, Gasoline | Using refinery technologies already in place, ease of adoption into current infrastructure | In early development stages. Refinery integration issues. | BIOCOUP Consortium (UOP, PNNL, NREL) | Development of process chain began in June 2006, concept development by 2011. Industrial demonstration may follow. |
| Fatty Acids to HydroCarbon - Hydrotreatment (FAHC - Hydrotreatment) | Seed Oil / Waste Oils / Animal Fats | Upgrading via hydrotreatment | Renewable Diesel | Using refinery technologies already in place, ease of adoption into current infrastructure | In early development stages. Refinery integration issues. | ConocoPhillips, Neste (UOP) | Pilot facilities in operation, others are planned for 2008 and beyond. |

1) Examples of some of the key players for each technology type. Universities and National Laboratories are shown in parenthesis.

2) High range of feedstock flexibility. Intermediate product from gasification can be used for broad slate of end-products (i.e. fuels, power, heat, etc.)

3) Gasification requires dried biomass. High level of syngas clean-up required.

Appendix B

Conversion Technology Pretreatment Matrix

Pretreatment requirements for selected biofuel conversion technologies

| Conversion Technology | Pretreatment Description | Required Moisture Content | Size | Source |
|--|---|---------------------------|---------|---|
| Grain to Ethanol - Dry Mill | Milling to a fine flour | -- | fine | <i>RFA 2005</i> |
| Grain to Ethanol - Wet Mill | Steeping | -- | -- | <i>RFA 2005</i> |
| Fatty Acid to Methyl Ester (FAME) | Filtering, dewatering | ~ 0% | -- | <i>Zhang et al. 2003a</i> |
| Lignocellulosics to Ethanol - Fermentation/Hydrolysis (1) | Cleaning (2), Sizing & Milling, <i>Dilute Acid Hydrolysis</i> | -- | 1-3 mm | <i>Hamelinck et al. 2005</i> |
| | Cleaning (2), Sizing, <i>Steam Explosion</i> | -- | 19 mm | <i>Hamelinck et al. 2005</i> |
| | Cleaning (2), Sizing, <i>Liquid Hot Water</i> | -- | 19 mm | <i>Hamelinck et al. 2005</i> |
| Lignocellulosics to Middle Distillates - Fischer Tropsch (3) | Sizing, Drying | 10-15 wt% | 6-13 mm | <i>Tijmensen et al. 2002</i> |
| Lignocellulosics to Gasoline - Upgrading/Pyrolysis | Sizing, Drying | 5-10 wt% | 2 mm | <i>Ringer, Putsche and Scahill 2006</i> |
| Fatty Acids to Hydrocarbon - Hydrotreatment | Combination of filtering, dewatering, acid washing, ion-exchange, desalting | ~ 0% | -- | <i>UOP 2005</i> |

1) This process has a drier to remove moisture from solid residuals before used for heat - usually available at MC of 60%, dried to 15%.

2) Cleaning is not required for all feedstocks. For example, de-barked forest residues do not generally need to be cleaned, while agricultural residues such as corn stover needs to be washed.

3) The required material size for gasification depends on the gasifier type and the feedstock. Woody biomass is typically sized to 1/4" minus, while agricultural residues are generally 1/2" minus, based on Antares experience.

Appendix C

Economic Analysis Factors

CDEAC Quantitative Work Group Economic Analysis Guidance

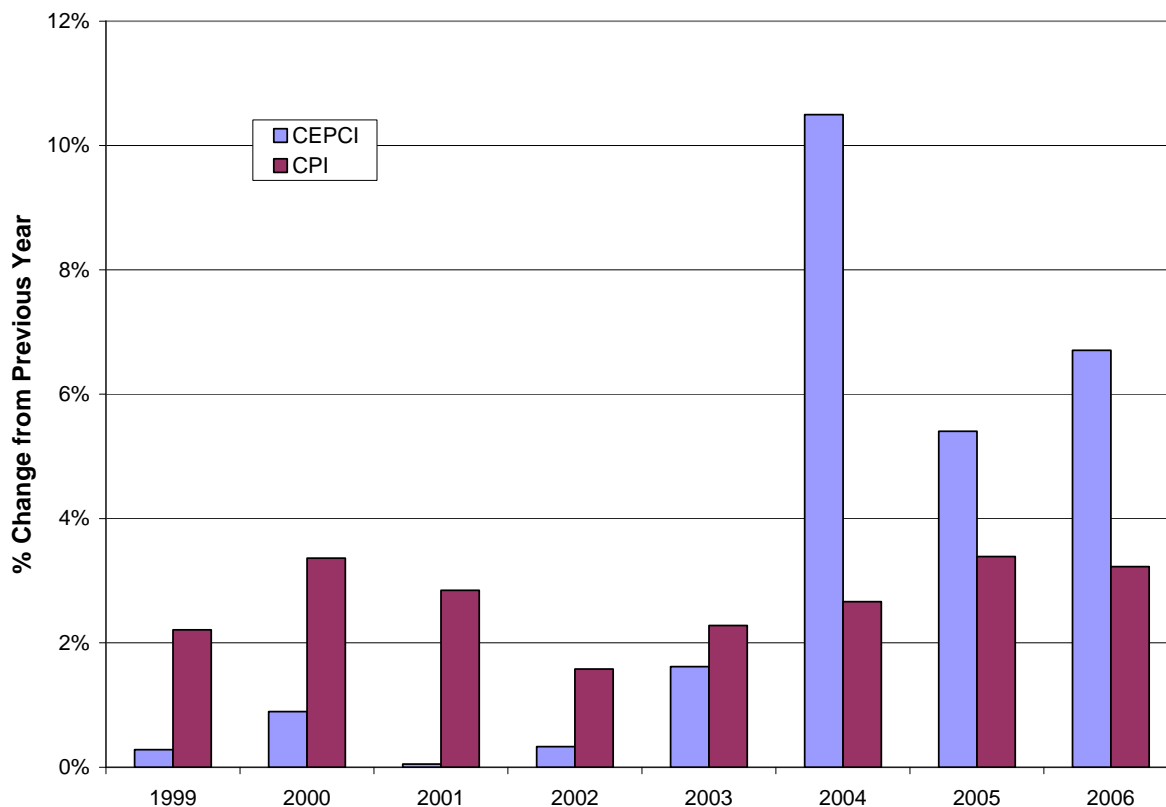
Factors for Levelized Production Cost Calculations

| | | |
|-------------------------------------|---------------|---|
| Equity Share | 45% | Yellow cells mark inputs recommended by Quantitative Work Group |
| Debt Share | 55% | |
| Nominal Return on Equity | 16.7% | |
| Nominal Return on Debt | 6.5% | |
| Federal Corporate Tax Rate | 35% | |
| State Corporate Tax Rate | 5% | |
| Effective Corporate Tax Rate | 38.2% | |
| Economic Book Life (years) | 25 | |
| Weighted Cost of Capital | 9.7% | |
| Present Value Depreciation Fraction | 0.778 | MACRS DB Depreciation |
| Capital Recovery Factor | 10.8% | |
| Fixed Charge Rate (FCR) | 12.26% | does not include production incentives, or additional property tax or insurance |

Nominal Non-Feedstock Levelized Production Cost (\$/gal) = $(FCR * \text{Total Capital Investment}(\$) + \text{Annual Operating Cost} (\$/\text{yr})) / \text{Biofuel Yield (gal/yr)}$

| MACRS - 5 Yr | | | Avg. State | |
|--------------|--------------|-------|----------------|---------------|
| Year | Depreciation | Fract | State | Corporate Tax |
| 1 | 0.2000 | 0.182 | AK | 5.2 |
| 2 | 0.3200 | 0.266 | AZ | 6.968 |
| 3 | 0.1920 | 0.145 | CA | 8.84 |
| 4 | 0.1152 | 0.079 | CO | 4.63 |
| 5 | 0.1152 | 0.072 | HI | 5.5 |
| 6 | 0.0576 | 0.033 | ID | 7.6 |
| | | | KS | 4 |
| | | | MT | 6.75 |
| | | | NE | 6.695 |
| | | | NM | 6.2 |
| | | | ND | 4.8 |
| | | | OR | 6.6 |
| | | | NV | 0 |
| | | | WA | 0 |
| | | | WY | 0 |
| | | | AVERAGE | 4.92 |

Comparison of Annual Change for CEPCI and CPI

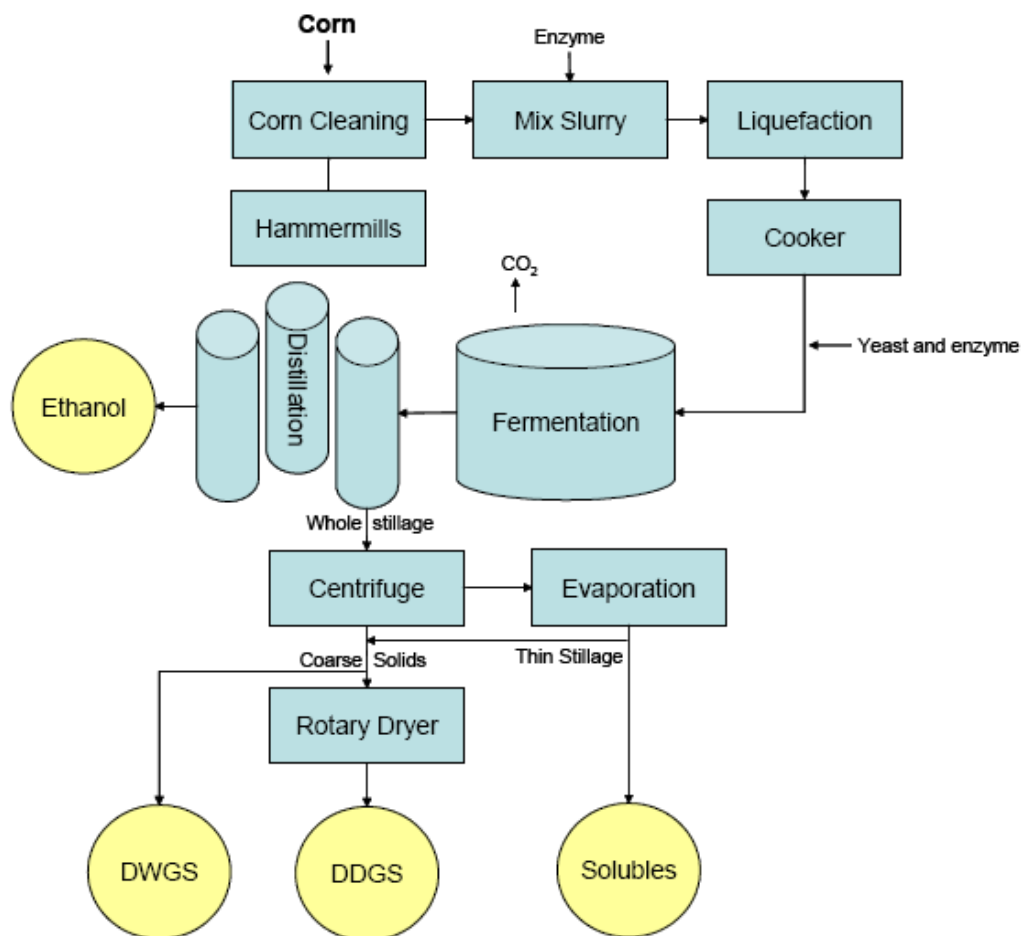


The graph above compares the annual change (from the previous year) for the Chemical Engineering Plant Cost Index (CEPCI) and the Consumer Price Index (CPI). In general, there is not a clear relationship between the CEPCI and the CPI. Furthermore, the CEPCI has seen large increases in recent years, particularly from 2003 to 2004. As a result of these increases, there is some concern within the biofuels industry about capital cost projections which will not adequately account for cost increases of materials used for chemical facilities (like steel). While this may indeed add significant costs to the capital, the same cost increases will apply to all fuel production industries. As such, the relative capital costs between the biofuel industries and oil refineries will remain the same.

Appendix D

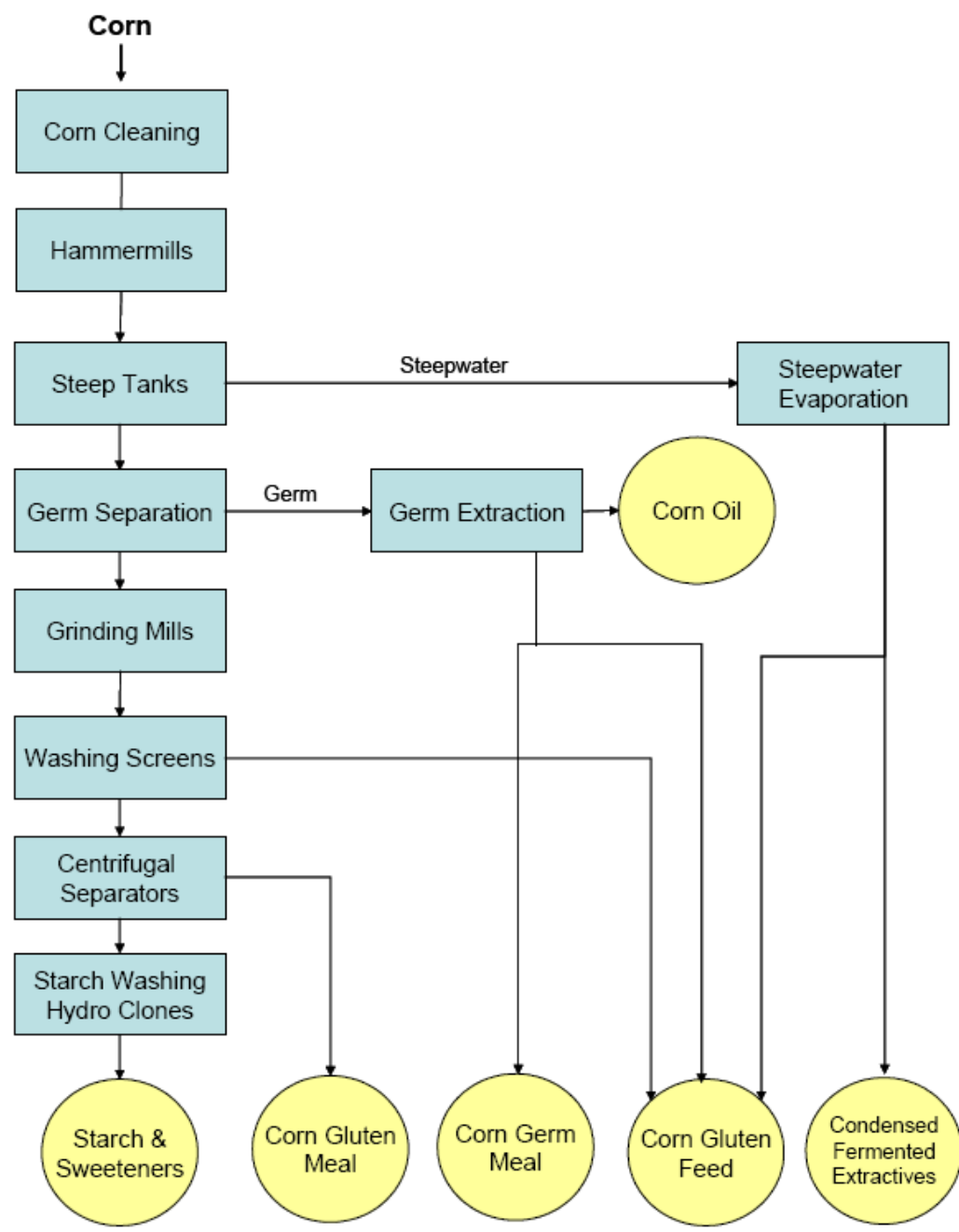
Additional Information - Grain to Ethanol Processes

Corn Dry-Grind Ethanol Milling Process Overview



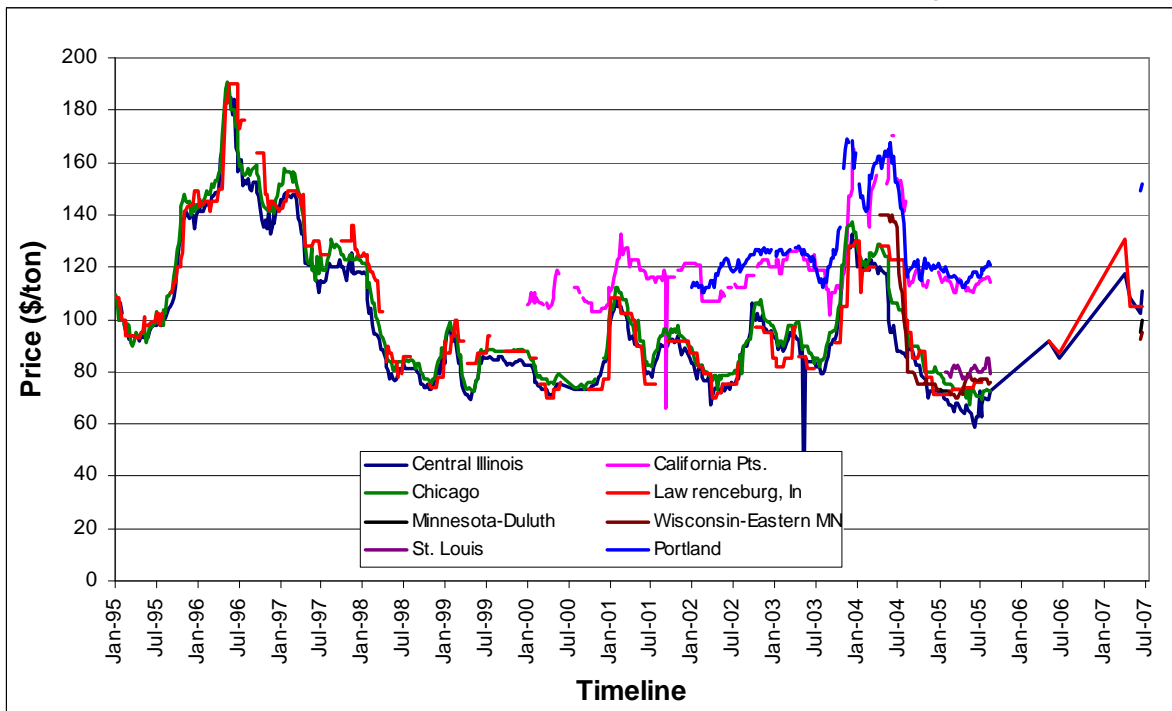
Source: Antares 2005a

Corn Wet Milling Process Overview



Source: Antares 2005a

DDGS 10-Year Historical Wholesale Price History⁶⁴



Source: Antares 2005b and USDA AMS 2007

As shown in the graph above, DDGS prices have fluctuated significantly over the past decade, ranging from \$60/ton to more than \$170/ton. Current DDGS prices are around \$90 to \$115 per ton, not including delivery costs.

⁶⁴ Note that the California and Portland prices are consistently higher than the other DDGS prices, as these numbers include the cost of rail delivery.

Appendix E

Oilseed Crop Feedstock Conversion to Oil

A typical method for conversion of oilseed feedstocks to oils is the n-hexane extraction process (Sheehan et al. 1998). Although alternate processes such as mechanical extraction are also available, hexane extraction has been selected for modeling in this analysis due to high oil yield and data availability.

CONVERSION

Table 21 shows the percentage oil in selected feedstocks. In general, the oil content is 20-40% of the total feedstock. The rest of the oilseed consists of hulls, moisture, meal, and other solids. For example, a typical break-down of soybeans includes about 20% oil, 7% hulls, 14-18% moisture, and 57% meal (approximate weight percentages from Sheehan et al. 1998). A large percentage of this non-oil product, including meal and hulls, is sold as animal feed. This is critical by-product and revenue source for the crushing/extraction process.

Table 21 – Oil content of various oilseed feedstocks

| Feedstock | Oil Content (Wt%) |
|-----------|-------------------|
| Soybean | 19% |
| Canola | 40% |
| Sunflower | 40% |
| Rapeseed | 40% |
| Safflower | 25% |

Based on data from Tyson et al. 2004

Sheehan et al. (1998) estimate that **oil extraction can recover 98.5% of the oil** in soybeans, including losses. The end product is pure degummed soybean oil ready for transesterification. For comparison, Tyson et al. (2004) suggest that the hexane extraction method can extract 99.5% of the oil from soybeans. However, this is the maximum extraction potential and does not fully adjust for losses.

ENERGY REQUIREMENTS

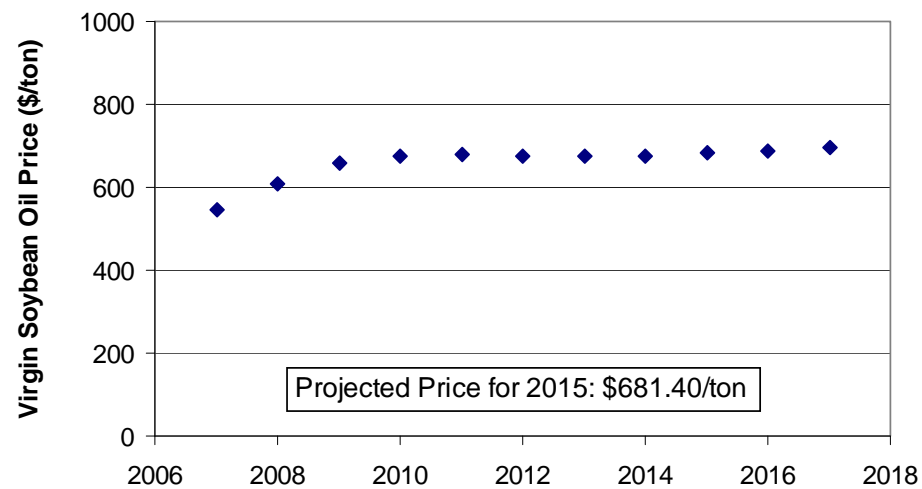
The energy requirements for the extraction process includes: 63.19 kWh/ton of electricity; 958.8 MMBtu/ton of natural gas; and 792.2 MMBtu/ton of steam (Sheehan et al. 1998).⁶⁵ Furthermore, extraction requires 4.04 lb hexane and 0.77 gallons of water per ton of feedstock processed (Sheehan et al. 1998).

PROJECTED PRICES

According to the Food and Agricultural Policy Research Institute (FAPRI) 2007 U.S. and World Agricultural Outlook, soybean oil is expected to be the largest source of feedstock for biodiesel in the mid-term. For the 2015 time period, FAPRI estimates that 75% of biodiesel will be produced from soy oil, 16% from canola oil, and the remainder produced from other fats and oils. The FAPRI 2007 price projections for virgin soybean oil and canola oil in the US are shown in Figure 16 and Figure 17, respectively. The projected soybean oil price is \$681.40/ton in 2015. The projected price for canola oil in 2015 is \$753.40/ton.

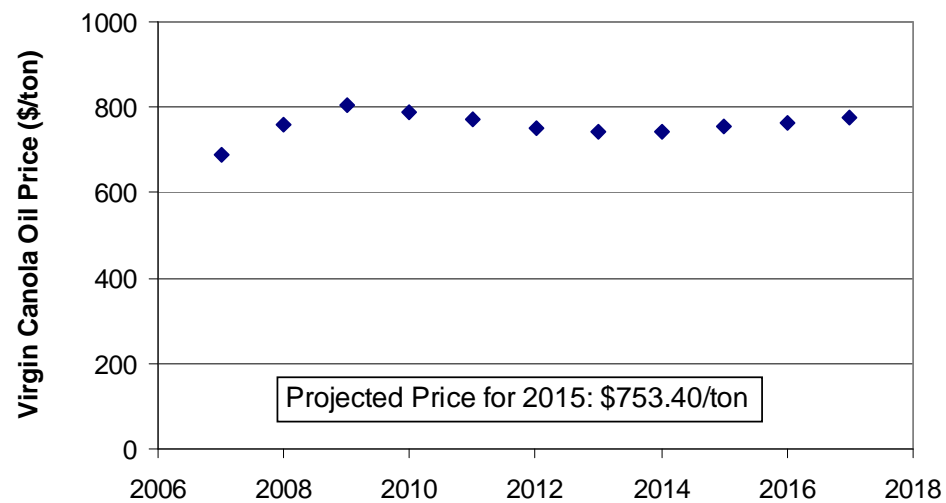
⁶⁵ All quantities given per ton of soybeans as received (i.e. with about 15% moisture). The natural gas is used for drying in the bean preparation step.

Figure 16 – Projected Soybean Oil Price



Data from FAPRI 2007

Figure 17 – Projected Canola Oil Price



Data from FAPRI 2007

Appendix F

Biomass Recovered from MSW

Biomass materials in the MSW stream include paper and cardboard, green waste, food waste, construction wood waste. Engineers at Taylor Recycling estimate that today's systems for culling biomass from the mixed stream will add about \$25 to \$30 per ton to the cost of biomass recovered from a mixed waste stream.⁶⁶ This cost covers the labor for hand picking clean materials or alternatively for rejecting visibly tainted materials, as well as O&M and capital recovery. Equipment for biomass recovery can include trommel screens, vibratory screens, air clarifiers, finger screen separators and shredders.

While much of the recovered biomass material can potentially be used for gasification-based processes like Fischer Tropsch (LCMD – FT), the same is not true for biochemical processes such as LCE Fermentation/ Hydrolysis. The LCE Fermentation/Hydrolysis process is particularly sensitive to feedstock composition quality and consistency as well as poisoning from contaminants, which limits the type of recovered materials that could be used. Based on discussions with industry experts, it has been determined that at least a portion of the recovered paper products and yard trimmings could potentially be used for the LCE Fermentation/Hydrolysis process, in addition to clean wood waste residues.

The compositions of the MSW materials applicable for the LCE Fermentation/Hydrolysis process are shown in Table 22. The paper composition is an average for newsprint and other paper. The yard waste is assumed to be a mix of leaves (20% by weight), grass (20%) and woody residues (60%), based on data from McKeever and Skog (2003). The yard waste composition is calculated as a weighted average of these materials.

Table 22 Composition of MSW Stream Components

| | Hemicellulose | Cellulose | Lignin |
|---|---------------|--------------|--------------|
| Paper, Average | 15.2% | 48.1% | 22.6% |
| <i>Paper</i> | 13% | 45% | 17% |
| <i>Newsprint paper</i> | 16% | 43% | 27% |
| <i>Newsprint paper (pre-consumer waste)</i> | 16% | 43% | 26% |
| <i>Newsprint</i> | 16% | 61% | 21% |
| Yard Waste | 31.8% | 20.0% | 20.6% |
| <i>Leaves, Average</i> | 20% | 24% | 17% |
| <i>Fallen leaves, oak</i> | 26% | 30% | 24% |
| <i>Mixed fresh leaves, wood</i> | 14% | 19% | 10% |
| <i>Grass - mowing waste</i> | 5% | 15% | 5% |
| <i>Wood, mixed</i> | 45% | 20% | 27% |
| <i>Poplar - hardwood</i> | 45% | 19% | 26% |
| <i>Pine - softwood</i> | 45% | 22% | 28% |

Data from (Energy Research Centre of the Netherlands, n.d.)

⁶⁶ Teleconference with Mark Paisley, Director of R&D for Taylor Recycling, on December 10, 2007

Appendix G

Conversion Technology Model Calculators